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Michigan Technological Univ., Houghton

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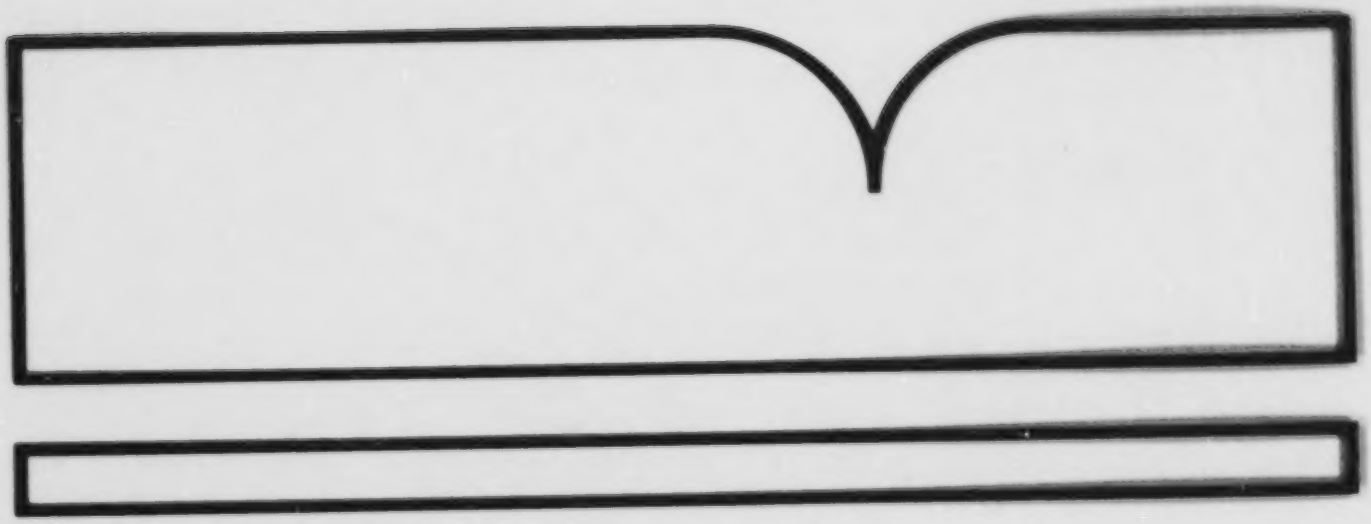

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16. ABSTRACT

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The results showed that the model and associated software package provide a useful capability to analyze, simulate, and predict oxygen utilization rates. It was possible to obtain reasonable agreement between the measured and simulated values of oxygen uptake rate, dissolved oxygen concentration and other process parameters at most of the plants studied. The key model parameters were the heterotrophic yield coefficient, heterotrophic decay constant, and autotrophic maximal specific growth rate constant.

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by

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Cooperative Research Agreement No. CR813162-01-2

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FOREWORD

Today's rapidly developing and changing technologies and industrial products and practices frequently carry with them the increased generation of materials that, if improperly dealt with, can threaten both public health and the environment. The U.S Environmental Protection Agency is charged by Congress with protecting the Nation's land, air, and water resources. Under a mandate of national environmental laws, the agency strives to formulate and implement actions leading to a compatible balance between human activities and the ability of natural systems to support and nurture life. These laws direct the EPA to perform research to define our environmental problems, measure the impacts, and search for solutions.

The Risk Reduction Engineering Laboratory is responsible for planning, implementing, and managing research, development, and demonstration programs to provide an authoritative, defensible engineering basis in support of the policies, programs and regulations of the EPA with respect to drinking water, wastewater, pesticides, toxic substances solid and hazardous wastes, and Superfund-related activities. This publication is one of the products of that research and provides a vital communication link between the researcher and the user community.

The research described in this report applies recent advances in activated sludge process modeling to the simulation of oxygen utilization in full scale activated sludge plants. This information is useful to engineers in the cost-effective design and operation of wastewater treatment systems.

E. Timothy Oppelt, Acting Director
Risk Reduction Engineering Laboratory

ABSTRACT

The objective of the research described in this report is to apply recent advances in activated sludge process modeling to the simulation of oxygen utilization rates in full scale activated sludge treatment plants. This is accomplished by calibrating the International Association for Water Pollution Research and Control (IAWPRC) Model and associated SSSP microcomputer software to operating data at six full scale activated sludge treatment plants. Field data were used to calibrate the key biological parameters contained in the model so that the oxygen utilization rates, dissolved oxygen concentrations, mixed liquor volatile suspended solids concentrations, and process performance simulated by the model matched the corresponding quantities observed in the treatment plants.

The results showed that the model and associated software package provide a useful capability to analyze, simulate, and predict oxygen utilization rates. It was possible to obtain reasonable agreement between the measured and simulated values of oxygen uptake rate, dissolved oxygen concentration and other process parameters at most of the plants studied. The key model parameters were the heterotrophic yield coefficient, heterotrophic decay constant, and autotrophic maximal specific growth rate constant.

This information is of value to engineers in the cost-effective design and operation of wastewater treatment systems because it provides a data base of applicable stoichiometric and kinetic model parameters which the engineer can utilize, with appropriate judgment, to simulate and predict the behavior of oxygen transfer systems in wastewater treatment.

This report was submitted in fulfillment of Cooperative Research Agreement CR813162-01-2 by Michigan Technological University under the partial sponsorship of the U.S. Environmental Protection Agency. This report covers a period from June 1, 1986 to June 30, 1988 and work was completed as of June 30, 1988.

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SYMBOLS AND NOMENCLATURE

DIMENSIONS

M = mass
L = length
t = time
T = temperature

SYMBOLS

b = decay coefficient, t^{-1}

Subscripts Used with b

a	denotes autotrophic decay coefficient
c	denotes conventional decay coefficient
h	denotes heterotrophic decay coefficient

C^*_∞ = average DO saturation concentration approached at infinite time

COD = chemical oxygen demand concentration, ML^{-3}
$SCOD_{eff}$ = soluble COD concentration in effluent, ML^{-3}
$SCOD_{inf}$ = soluble COD concentration in influent, ML^{-3}
$TCOD_{inf}$ = total COD concentration in influent, ML^{-3}

DO = dissolved oxygen concentration, ML^{-3}

f_O = fraction of biomass yielding particulate products upon oxidative decay

f_P = fraction of biomass yielding particulate products upon hydrolytic decay

i_{XB} = mass N/mass COD in biomass

i_{XP} = mass N/mass COD in products from biomass

K = half saturation constant, ML^{-3}

Subscripts Used with K

OA	denotes autotrophic constant for dissolved oxygen
OH	denotes heterotrophic constant for dissolved oxygen
NH	denotes autotrophic constant for ammonia N

NO denotes heterotrophic constant for nitrate N
S denotes constant for soluble substrate, COD
X denotes hydrolysis constant, dimensionless

K_A = heterotrophic ammonification rate, $L^3M^{-1}t^{-1}$

K_La = volumetric mass transfer coefficient for oxygen, t^{-1}

k_N = heterotrophic hydrolysis rate, t^{-1}

MGD = million gallons per day

MLVSS = mixed liquor volatile suspended solids concentration, ML^{-3}

N_{TA} = amount of nitrogen available for nitrification, M/t

N_{TN} = amount of total nitrogen converted to nitrate, M/t

N_{NO} = amount of nitrate in the effluent, M/t

N_{DN} = amount of nitrogen denitrified, M/t

OTE = oxygen transfer efficiency, dimensionless

OUR = oxygen uptake rate, $ML^{-3}t^{-1}$

Q = flow rate entering and leaving the process, L^3t^{-1}

Q_W = waste sludge flow, L^3t^{-1}

R_n = nitrogenous oxygen demand, M/t

R_t = total carbonaceous and nitrogenous oxygen utilization rate, M/t

R_C = rate of oxygen utilization by carbon oxidation processes, M/t

r_o = specific oxygen utilization rate, t^{-1}

S = soluble material concentration, ML^{-3}

Subscripts Used with S

C denotes organics in conventional model, COD
C1 denotes influent organics in conventional model, COD
C2 denotes effluent organics in conventional model, COD
AIK denotes alkalinity, moles/liter
I denotes inert organic matter, COD
O denotes dissolved oxygen, negative COD
O1 denotes influent dissolved oxygen, negative COD
O2 denotes effluent dissolved oxygen, negative COD
NH denotes ammonia nitrogen
S denotes readily biodegradable substrate, COD

NO denotes nitrate plus nitrite nitrogen
ND denotes biodegradable organic nitrogen

SRT = solids retention time, t

TKN = total Kjeldahl nitrogen, M/L^3
TON_{inf} = total organic nitrogen in influent, ML^{-3}

U = specific substrate utilization rate, t^{-1}

V = reactor volume

W_{O2} = mass rate of oxygen applied, M/t

X = particulate material concentration, ML^{-3}

Subscripts Used with X
B denotes biomass in conventional model, COD
B1 denotes influent biomass, COD
B2 denotes effluent biomass, COD
BA denotes active autotrophic biomass, COD
BH denotes active heterotrophic biomass, COD
P denotes decay products, COD
I denotes inert organic matter concentration, COD
I1 denotes influent inert organic matter, COD
ND denotes biodegradable organic nitrogen, N
P denotes products arising from oxidative decay, COD
S denotes slowly biodegradable substrate, COD

Y = yield coefficient, bimass COD/substrate COD

Subscripts Used with Y
a denotes autotrophic yield coefficient
H denotes heterotrophic yield coefficient

η_H = correction factor for anoxic hydrolysis

θ_C = SRT at steady state conditions, t

η_G = correction factor for anoxic growth of heterotrophs

μ = specific growth rate, t^{-1}

Subscripts Used with μ
c denotes conventional specific growth rate
cm denotes conventional maximal specific growth rate
hm denotes autotrophic specific growth rate
m denotes maximal specific growth rate
am denotes autotrophic specific growth rate

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This study utilized off-gas and plant operating data collected by other investigators working on the EPA - ASCE Fine Bubble Diffused Aeration Design Manual Project. Such data and information contributed by David Redmon of Ewing Engineering, William Boyle of the University of Wisconsin, James Marx of Donohue & Associates, and Read Warriner of the Milwaukee Metropolitan Sewerage Commission are gratefully acknowledged.

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SECTION 1

INTRODUCTION

The activated sludge process is the most widely used method for secondary wastewater treatment in the United States, and its popularity is increasing. Provision of oxygen to the active organisms through aeration is the most energy intensive aspect of activated sludge process operation and consumes 60 to 80% of the total energy requirements in wastewater treatment. Because of this, approximately 1.75 million horsepower of aeration equipment is currently installed in wastewater treatment plants in the United States and Canada. The energy cost for operating this equipment amounts to more than $600 million per year. It is estimated that, through improvements in design and operation of these aeration systems, savings of more than $100 million per year could easily be achieved. Furthermore, many municipalities are replacing older, less efficient aeration systems with more energy-efficient fine bubble diffused aeration systems, and additional savings in repair and replacement costs could be realized.

For the efficient design and operation of aeration systems it is necessary that the oxygen demands of the biological system and the oxygen transfer capability of the aeration equipment be accurately predicted. Over the past few years, significant progress has been made in measuring and modeling the oxygen transfer capability of aeration equipment (ASCE, 1984) (Brown and Baillod, 1982)(Mueller and Boyle, 1988) under clean water and process conditions. However, the progress in modeling equipment performance capability has not been matched by improvements in modeling, simulating, and designing for oxygen transfer requirements in full scale activated sludge plants. The methods commonly used at the present time are adequate only for estimation of the average oxygen utilization rate, and are of limited practical usefulness in design of efficient new and replacement aeration systems. With these conventional methods, the designer must make an educated guess of the spatial and temporal variation in oxygen utilization rates, and this can lead to over-design or under-design of an aeration system. An under-designed system produces a poorly treated effluent, whereas an over-designed system has high initial and operating costs.

The objective of the research described in this report is to apply recent advances in activated sludge process modeling (Grady et al., 1986) (Bidstrup and Grady, 1988) to the simulation of oxygen uptake rates in full scale activated sludge treatment plants. This is accomplished by calibrating the International Association for Water Pollution Research and Control (IAWPRC) Model and associated SSSP software to operating data at six full scale activated sludge treatment plants.

This information is of value to engineers in the cost-effective design and operation of wastewater treatment systems because it provides a data base of applicable stoichiometric and kinetic model parameters which the engineer can utilize, with appropriate judgment, to simulate and predict the behavior of oxygen transfer systems in wastewater treatment.

SECTION 2

CONCLUSIONS

Field studies were conducted at six municipally owned and operated activated sludge wastewater treatment plants in order to assess and enhance the usefulness of a mathematical model and associated microcomputer software package for simulating the oxygen utilization rate (OUR) of the activated sludge process. The results of the field studies were used to calibrate the key biological parameters contained in the model so that the oxygen utilization rates, dissolved oxygen (DO) concentrations, mixed liquor volatile suspended solids concentrations, and process performance simulated by the model matched the corresponding quantities observed in the treatment plants. Based on the results of this study, it can be concluded that:

1. The International Association for Water Pollution Research and Control (IAWPRC) Model and related SSSP (Simulation of Single-Sludge Processes for Carbon Oxidation, Nitrification and Denitrification) microcomputer software package provide a useful capability to analyze and simulate the average oxygen utilization rate in municipal wastewater treatment plants.

2. It was possible to obtain reasonable agreement between the average measured and steady-state simulated values of oxygen uptake rate, dissolved oxygen concentration, mixed liquor volatile suspended solids, effluent ammonia, and effluent nitrate concentrations at most of the plants studied. This agreement was achieved by adjusting or calibrating only two of three key model parameters while keeping the other 17 model parameters at their adjusted default values. The key model parameters were the heterotrophic yield coefficient, heterotrophic decay constant, and autotrophic maximal specific growth rate constant.

3. When provided with a realistic value of the process water volumetric mass transfer coefficient (K_La), the calibrated model and software were normally able to simulate the spatial and temporal range of dissolved oxygen concentrations and oxygen utilization rates observed in the operating treatment plants. Instances in which lack of agreement between simulated and observed values occurred could be possibly explained by causes other than model and/or software inadequacy.

4. Under conditions of low DO concentration (below 1 to 2 mg/L) *in vitro* OUR values indicated by the traditional BOD bottle method tended to be greater than the *in situ* OUR values determined from off gas analyses.

These differences were most evident in the inlet sections of long aeration tanks where OUR values were high and DO values were low.

5. Comparison of the IAWPRC model and conventional approaches to estimate average process oxygen utilization rates showed that, for given values of the yield and decay coefficients, the two approaches agreed reasonably well. In this study, the IAWPRC model was calibrated to the data and, therefore, was judged to give the better estimates of the process utilization rates. Moreover, the IAWPRC model was judged to be advantageous because of its ability to simulate nitrogenous oxygen demand.

SECTION 3

RECOMMENDATIONS

This study relied upon adjustment or calibration of key parameters to show that the IAWPRC Model could be applied to analyze, simulate, and predict oxygen utilization and other activated sludge process characteristics. Such information is valuable because it provides a data base of key parameters from which engineers can draw information for process design and operation. However, this data base should be expanded to include information from more than the six plants of this study. Moreover, it would be of particular value to demonstrate the applicability of techniques for key parameter estimation based on direct measurements.

Accordingly, it is recommended that additional in-depth studies applying the IAWPRC Model and associated SSSP Software to well controlled full-scale wastewater treatment plants be conducted. The plants should be selected so that the process train studied treats a sizable fraction (at least one-third) of the wastewater flow monitored, and intensive data collection should be carried out for a period of about two weeks at each plant. The intensive data collection should include the items measured in the 24 hour plant studies described in this report, plus periodic off gas measurements of K_La,and repeated direct measurements of the key model stoichiometric and kinetic parameters and wastewater characteristics.

It is also recommended that design engineers be encouraged to apply the IAWPRC Model and associated SSSP Software to activated sludge process design. Although other improved models and software might be developed in the future, this model and software are available now and are easily usable by any design engineer. With the aid of engineering judgment and the parameter data base given in this report, this approach will facilitate more economical and effective design and operation of wastewater treatment plants.

SECTION 4

MODELS APPLIED TO OXYGEN UTILIZATION IN BIOLOGICAL WASTE TREATMENT SYSTEMS

REQUISITES FOR BIOLOGICAL PROCESS MODELS

Design and operation of biological waste treatment systems are aided by application of predictive models which can mimic the performance of these systems. Because biological waste treatment systems are inherently complex, it is necessary that the model be somewhat simpler than the actual system. Over the past several years, various conceptual models ranging from single substrate, single biomass aerobic models (Herbert, 1956) to multiple substrate, multiple biomass anaerobic models (Rozzi et al., 1985) have been applied to biological waste treatment.

Successful modeling of a system requires:

1. A substantially correct conceptual representation of the biological processes, stoichiometry, kinetics, and pathways involved.

2. Definition of the physical system, i.e. tanks, pipes, flow rates, in sufficient detail so that material balance equations can be written for each component of interest.

3. An analytical or numerical solution of the material balance equations.

4. Verification of the model principles and calibration of the model parameters by application of the model to real treatment systems.

Early modeling efforts were hampered by a multiplicity of conceptual representations and a lack of available computing technology for model solution. In 1983, the International Association on Water Pollution Research and Control (IAWPRC) established a task group on "Mathematical Modeling for Design and Operation of Biological Waste-water Treatment". The assignment of this group was to develop a consensus model applicable to an activated sludge system performing simultaneous carbon oxidation, nitrification and denitrification. The model which resulted from this effort is termed the IAWPRC Model and is described by Grady et al. (1986), Henze et al. (1987), and the IAWPRC Task Group (1986). Basic features of this model are discussed in the following section.

In 1987, Bidstrup and Grady (1987) developed the SSSP (Simulation of Single-Sludge Processes for Carbon Oxidation, Nitrification and Denitrification) computer software package based on the IAWPRC Model. This

software, written for the IBM Personal Computer or compatible machines performs steady-state and dynamic simulations of activated sludge systems based on the IAWPRC Model. The program is versatile in that it allows the user to define the system configuration by using up to nine completely mixed reactors in series. Additional flexibility is provided by the capability to define the influent addition, return sludge recycle, and mixed liquor recirculation flow diagrams between the various reactors.

The work of the IAWPRC Task Group, coupled with the SSSP software package, satisfied the first three Requirements for successful modeling of a biological waste treatment system. One of the objectives of this research effort is to help satisfy Requirement 4, i.e. to verify and calibrate the IAWPRC model by applying it to simulate oxygen utilization in full-scale activated sludge plants.

OXYGEN REQUIREMENTS IN ACTIVATED SLUDGE SYSTEMS

Oxygen requirements in activated sludge systems arise from:

* Biological oxidation of organic carbon
* Biological oxidation of nitrogen
* Chemical oxidation of inorganic substances

The rates of oxygen utilization by each of these processes are functions of the process influent load and operating conditions. Efficient design of oxygenation or aeration systems requires reliable estimates of the average, minimum and maximal total oxygen utilization rates. In addition, information on the spatial variation of oxygen utilization rate within the process must be known. Also, aeration frequently is relied upon to mix the fluid to promote mass transfer and to keep the biomass particles suspended. In lightly loaded activated sludge or aerated lagoon systems, the mixing requirement may control.

Empirical Approaches

In the past, various empirical and rule-of-thumb approaches have been used to estimate oxygen requirements for activated sludge systems:

* 1,500 cubic feet of air per pound of 5-day BOD applied
* 1 pound of oxygen transferred per pound of 5-day BOD applied (Great Lakes - Upper Mississippi River Board of State Sanitary Engineers, 1978)
* 0.5 to 2.0 cubic feet of air per gallon of sewage (Fair and Geyer, 1954)
* 500 to 700 cubic feet of air per pound of 5-day BOD removed with at least 3 cfm per foot of tank length for mixing (Water Pollution Control Federation, 1977)

Conventional Model Applied to Heterotrophic Oxygen Utilization

More recently, rational approaches based on the overall process oxygen,

substrate, and biomass balances have been developed to estimate carbonaceous oxygen demand (Grady and Lim, 1980). The following development illustrates the use of a matrix table (Grady et al. 1986) to formulate material balances for the conventional carbonaceous oxygen utilization model described in Figure 1.

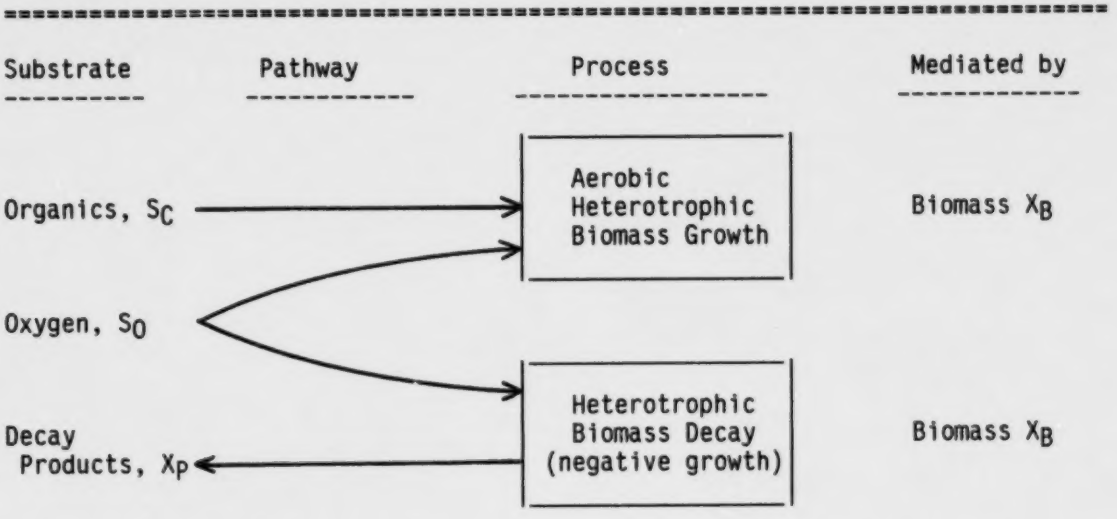

Figure 1. Substrates, pathways and processes included in the conventional carbonaceous oxygen utilization model.

It should be noted that, throughout this report, concentrations of carbonaceous substrates and biomass are expressed as total carbonaceous biochemical oxygen demand (TCBOD). This is used synonymously with biodegradable chemical oxygen demand (BCOD). In the literature, this is often referred to as ultimate biochemical oxygen demand or as total biochemical oxygen demand. Biomass volatile suspended solids are assumed to have the composition $C_5H_7NO_2$, and are converted to oxygen equivalent by the factor 1.41 g oxygen/gram VSS.

Explanation of Matrix Table for Model Representation--

For complex models in which several processes operate to transform several components, it is convenient to use a matrix table to represent the kinetics and stoichiometry of the model. Table 1 is an example of a matrix table for the model described in Figure 1. The model considers only aerobic heterotrophic processes in biological waste treatment. The components of importance in the model are listed by symbol across the top of the table, and their definitions are given at the bottom of the corresponding columns.

This model considers only two processes, aerobic growth of heterotrophic biomass, and oxidative decay of heterotrophic biomass. The decay process is termed "oxidative" because it utilizes oxygen. It is

Table 1. Kinetics and Stoichrometry for Conventional Carbonaceous Oxygen Utilization Model

Component	1 S_I	2 S_C	3 X_I	5 X_B	7 X_P	8 S_O	Process Rate
Aerobic Growth of Heterotrophic Biomass		$1 - \frac{1}{Y_H}$		+1		$-\frac{1-Y_H}{Y_H}$	$\mu_c X_B = \mu_{cm} \left(\frac{S_S}{K_S + S_S}\right) \left(\frac{S_O}{K_{OH} + S_O}\right) X_B$
Heterotrophic Biomass Decay				-1	f_o	$-(1-f_o)$	$b_c X_B$
	Dissolved Oxygen Concentration, negative COD	Particulate Products Arising from Oxidative Decay, Concentration, COD	Biomass Concentration, COD	Particulate Inert Organic Matter Concentration, COD	Soluble Substrate Conc., COD	Soluble Inert Organic Matter Concentration, COD	

Stoichrometric Parameters (2)

Yield Coeff: Y_H

Fraction of Biomass Yielding Particulate Products upon Oxidative Decay: f_o

Kinetic Parameters (4)

Conventional Maximum Specific Growth Rate: μ_{cm}

Half Velocity, Constants for Substrate and Oxygen: K_S, K_{OH}

Conventional Oxidative Decay Rate: b_c

important to distinguish this decay process from the hydrolytic decay process incorporated into the IAWPRC Model. The hydrolytic process uses no oxygen, and because of this, the magnitude of the hydrolytic decay coefficient is roughly 2.5 times the oxidative decay coefficient. Reasons for this difference are discussed later. The model assumes that each unit of biomass that undergoes decay is partially oxidized and partially converted to particulate products which are resistant to further decay. The observed fraction of biomass converted to resistant particulate products, X_P, has been reported to be 0.2 (IAWPRC Task Group, 1986).

The processes are listed down the extreme left column, and the corresponding rate equations are given in the extreme right column. It is important to realize that the process rate equations are expressed in dimensions of (biomass COD)/(time)/(volume). The stoichiometric coefficients listed in the body of the table have dimensions of (mass of component COD)/(mass of biomass COD). Thus, multiplication of a stoichiometric coefficient times the process rate equation gives the component transformation rate, as (mass of component COD)/(time)/(volume), for that process. Therefore, proceeding down a column and summing the products of the stoichiometric coefficients times the process rate equations will give the rate of component transformation for use in component material balances. A typical component material balance can be stated as:

$$\begin{array}{c}\text{Rate of}\\ \text{Component In}\\ \text{By Flow}\end{array} - \begin{array}{c}\text{Rate of}\\ \text{Component Out}\\ \text{By Flow}\end{array} + \begin{array}{c}\text{Rate of}\\ \text{Component}\\ \text{Production}\end{array} = \begin{array}{c}\text{Rate of}\\ \text{Component}\\ \text{Accumulation}\end{array} \qquad (1)$$

The rate of oxygen production, for example, is given by summing the products of the coefficients listed under S_O times their corresponding rate equations.

$$\begin{array}{c}\text{Rate of}\\ \text{Oxygen}\\ \text{Utilization}\end{array} = - \begin{array}{c}\text{Rate of}\\ \text{Oxygen}\\ \text{Production}\end{array} = \frac{1-Y_H}{Y_H}\,\mu_C V + (1-f_O)\, b_C X_B V \qquad (2)$$

where: V = reactor volume, and the other symbols are defined in Table 1.

Stoichiometric Coefficients--

By definition, the yield coefficient relates the biomass growth rate to the substrate utilization rate.

$$Y_H = \frac{\text{Specific Biomass Growth Rate}}{\text{Specific Substrate Utilization Rate}} = \frac{\mu_C}{U} \qquad (3)$$

where U = specific substrate utilization rate, t^{-1}

Hence, the appropriate stoichiometric coefficient by which to multiply the biomass growth rate expression to obtain substrate utilization rate is the negative reciprocal of the yield coefficient, i.e. $-U/\mu_C = -1/Y_H$. This factor, therefore, appears in row 1, column 2 of Table 1.

When substrate and biomass are both expressed as total carbonaceous biochemical oxygen demand TCBOD, or biodegradable COD, the rate of substrate TCBOD utilized for growth is equal to the rate of oxygen utilization for biomass growth plus the TCBOD of the biomass produced, or

$$U = r_O + \mu_C \tag{4}$$

where r_O = specific oxygen utilization rate, $1/t^{-1}$

The appropriate stoichiometric coefficient by which to multiply the biomass rate expressions to obtain oxygen utilization rate is $-r_O/\mu_C$. Combination of Equations 3 and 4 gives this coefficient as

$$-r_O/\mu_C = -(U/\mu_C - 1) = -(1-Y_H)/Y_H \tag{5}$$

Use of Matrix Table for Formulation of Component Material Balances

Consider the input to and output from the activated sludge process shown in Figure 2.

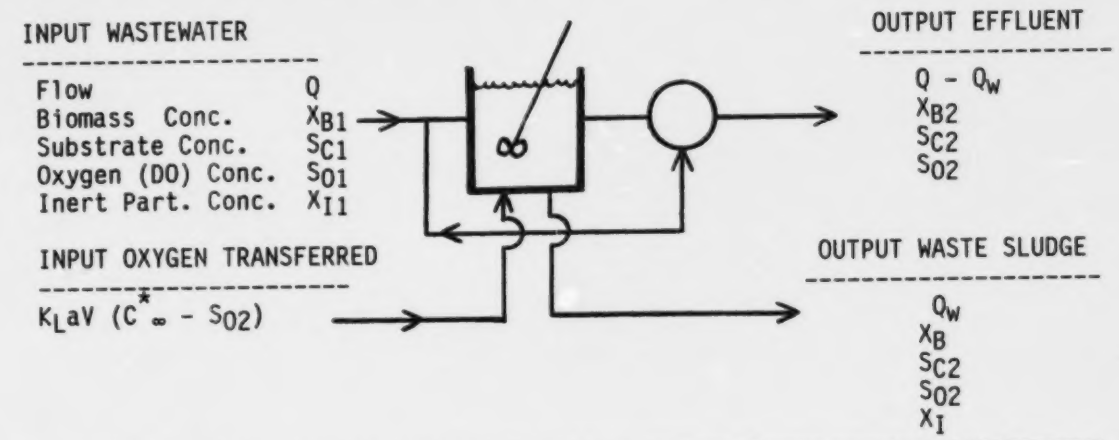

Figure 2. Activated sludge process material balance.

The process material balances can be written with the aid of the matrix shown in Table 1.

Biomass Balance

$$QX_{B1} - (Q-Q_W)X_{B2} - Q_W X_B + (\mu-b)X_B V = \frac{d}{dt}(VX_B) \tag{6}$$

Substrate Balance

$$QS_{C1} - Q_W S_{C2} - (Q-Q_W)S_{C2} - \frac{1}{Y_H}\mu_C X_B V = \frac{d}{dt}(VS_{C2}) \quad (7)$$

Oxygen Balance

$$QS_{O1} - Q_W S_{O2} - (Q-Q_W)S_{O2} - \left(\frac{1-Y_H}{Y_H}\right)\mu_C X_B V - b_C X_B V (1-f_O) + K_L a\, V(C^*_\infty - S_{O2}) = \frac{d}{dt}(VS_{O2}) \quad (8)$$

where, C^*_∞ = DO saturation concentration approached at infinite time in the unsteady state oxygen transfer test (ASCE, 1984).

Particulate Products Balance

$$Q(0) - Q_w X_P - (Q-Q_w)\, X_{P2} + f_O b_C X_B V = \frac{d}{dt}(VX_P) \quad (9)$$

Inert Particulates Balance

$$QX_{I1} - Q_w X_I - (Q - Q_w)\, X_{I2} - 0 = \frac{d}{dt}(VX_I) \quad (10)$$

From Table 1, the rate of oxygen utilization, R_C, is defined by,

$$R_C = \left(\frac{1-Y_H}{Y_H}\right)\mu_C X_B V - (1-f_O) b_C X_B V \quad (11)$$

Consideration of the relative magnitudes of the terms in the oxygen balance shows that the dissolved oxygen advection terms (the first three in Equation 8) are small compared to the other terms and can be neglected. Likewise, it is assumed that the particulate products X_{P2}, and inert particulate, X_{I2}, in the effluent can be approximated by zero. Steady state conditions are assumed so that all derivatives with respect to time are zero. With these simplifications, the biomass, substrate, oxygen, and particulate products balances can be combined with Equation 11 to express the oxygen utilization rate as,

$$R_C = Q(S_{C1} - S_{C2})\left(\frac{1 + b_C\theta_C - Y_H - f_O Y_H b_C \theta_C}{1 + b_C\theta_C}\right) \quad (12)$$

where:

R_C = mass of oxygen required per unit time for the carbon oxidation processes, M/t

Q = flow rate entering and leaving the process, L^3/t

S_{C1}, S_{C2} = total carbonaceous oxygen demand (ultimate BOD) of the flows entering and leaving the process, M/L^3

b_C = conventional decay coefficient for oxidative decay of biomass, 1/t

Y_H = yield coefficient, mass of volatile suspended solids COD produced per unit mass of carbonaceous oxygen demand utilized

θ_C = solids retention time (SRT) at steady state conditions, defined as the biomass in inventory divided by the rate of biomass wasting, t

f_O = fraction of biomass yielding nonbiodegradable particulate products upon oxidative decay, generally taken as 0.2

The decay coefficient, b_C, and yield coefficient, Y_H, are biological parameters characteristic of the biomass itself, whereas the solids retention time, θ_C, is the key parameter which controls the process operation and performance. Typical values of solids retention time range from 3 to 12 days. Numerical values of Y depend on the units in which biomass and substrate are expressed. Typical values for Y range from 0.45 to 0.7 mass of VSS COD produced per unit mass of feed COD utilized. Typical values of the oxidative decay coefficient, b_C, range from 0.04 to 0.4 per day (Grady and Lim, 1980). A variation of this model (Grady and Lim, 1980) neglects the particulate products component and becomes equivalent to this model when f_O is taken as zero.

In the general process analysis and design solution for this problem, (a single completely mixed aeration tank with recycle) another biomass balance is written on the reactor itself to include the recycle stream, and use is made of the kinetic relationship for μ_C shown in Table 1. For fixed values of the input quantities, output quantities (S_{C2} is very small but is not fixed), stoichiometric parameters, and kinetic parameters, the design problem has four degrees of freedom. Typically, the biomass concentration, X_B, the solids retention time, θ_C, the reactor dissolved oxygen concentration, S_{O2}, and the biomass concentration in the recycled sludge are chosen as design variables.

This model may also be applied to a more complicated flow diagram such as tanks in series, contact stabilization, or step feed. However, the material balance equations become more difficult to represent and solve. Furthermore, there is little reason to do so as the SSSP software package already supplies solutions for a more realistic model applied to a wide variety of process configurations.

Conventional Model for Nitrogenous Oxygen Utilization

The overall nitrogenous oxygen utilization rate can be calculated based on (EPA 1975):

1. The amount of nitrogen available for nitrification, N_{TA}, mass/time, determined as the total Kjeldahl nitrogen (TKN) in the influent less the nitrogen in the waste sludge.

2. The amount of total nitrogen converted to nitrate, N_{TN}, mass/time, determined by subtracting the TKN of the effluent from the total available nitrogen.

3. The amount of nitrate in the effluent, N_{NO},mass/time, measured or estimated.

4. The amount of nitrogen denitrified, N_{DN}, mass/time, determined by subtracting the nitrate in the effluent from the total nitrogen converted to NO_3.

The nitrogenous oxygen demand is then determined based on the stoichiometric requirements that:

* 1 mg of ammonia or Kjeldahl nitrogen requires 4.57 mg of oxygen for conversion to nitrate, and

* denitrification of 1 mg of nitrate nitrogen provides an equivalent oxygen credit of 2.86 mg.

Therefore, the nitrogenous oxygen demand, R_n, can be expressed as:

$$R_n = 4.57\ N_{TN} - 2.86\ N_{DN} \qquad (13)$$

And the total carbonaceous and nitrogenous oxygen utilization rate, R_t is then given by,

$$R_t = R_C + R_n \qquad (14)$$

Evaluation of Conventional Models for Oxygen Utilization

The conventional modeling approach described above is useful and certainly represents a great improvement over the empirical methods. However, it has the following disadvantages:

1. The model itself is deficient in that it does not adequately distinguish between slowly and readily degradable substrates; nor does it allow for replenishment of soluble substrate through hydrolysis.

2. The conventional nitrogenous model requires the user to estimate the amount of nitrification and denitrification attained by estimating the TKN and nitrate concentrations in the effluent. It does not provide any capability to simulate or predict these concentrations based on wastewater or process parameters. Consequently, the nitrogenous oxygen demand calculations are only as precise as the estimates.

These disadvantages are largely eliminated by the IAWPRC Model and the SSSP Software Package.

International Association for Water Pollution Research and Control (IAWPRC) Model Applied to Heterotrophic and Autotrophic Oxygen Utilization

The substrates, pathways, and processes included in the IAWPRC Model are

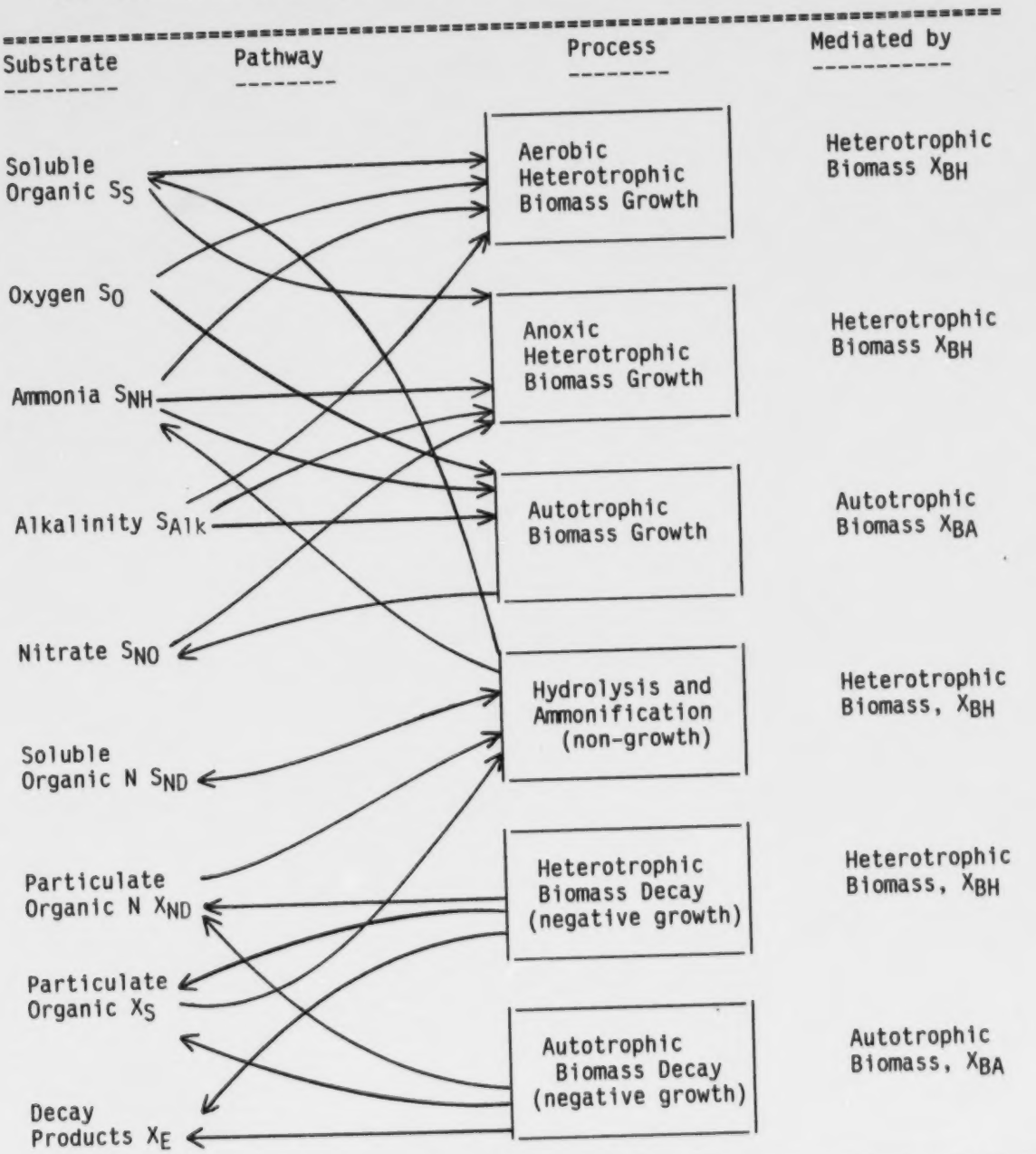

Figure 3. Substrates, pathways and processes included in the IAWPRC model.

described in Figure 3. Comparison of Figure 3 with Figure 1 shows the comprehensive nature of the IAWPRC Model. The model contains 9 substrates which are involved in 8 processes mediated by 2 active biomass fractions. The increase in complexity of this model over the conventional carbonaceous model is caused by:

1. Inclusion of autotrophic nitrification and other nitrogenous pathways. This adds four processes (autotrophic growth, autotrophic decay, ammonification, and hydrolysis of organic nitrogen) and five components (autotrophic biomass, nitrate, ammonia, particulate organic nitrogen, and soluble organic nitrogen).

2. Recognition of two classes of carbonaceous substrate, readily degradable (soluble) organics, and slowly degradable (particulate) organics. This adds one additional component and one additional process.

3. Inclusion of denitrification by anoxic growth of heterotrophs. This adds one additional process.

Table 2 shows the component/process matrix for the IAWPRC Model. The corresponding process rate equations are given in Table 3, and the definitions of process symbols are given in Table 4. Table 5 summarizes the kinetic and stoichiometric parameters contained in the model.

The matrix table is straight-forward and can be applied to formulate the reaction terms for system material balances in the same fashion that Table 1 was applied to formulate material balances. The stoichiometric coefficients listed for the heterotrophic growth processes are similar to those explained earlier for Table 1. The factors 4.57 and 2.86, respectively, represent the nitrification demand per unit ammonia nitrogen and the denitrification credit per unit of nitrate nitrogen. They appear because both ammonia and nitrate are expressed as concentration of nitrogen rather than as COD, and because the autotrophic yield is expressed as autotrophic biomass COD produced/mass of nitrogen utilized.

An important difference between the IAWPRC and conventional carbonaceous models arises because of the hydrolytic (as opposed to oxidative) nature of the decay processes built into the IAWPRC Model. Note that in the conventional model described by Table 1, the decay process utilizes oxygen to oxidize $(1-f_D)$ fraction of biomass COD and converts f_D fraction of biomass into resistant particulate products. However, in the IAWPRC Model, the decay processes use no oxygen. Instead, $(1-f_p)$ fraction of biomass is hydrolyzed to slowly degradable substrate, X_S, which, in turn, is hydrolyzed to readily degradable substrate, S_S. Utilization of oxygen is accounted for only by aerobic growth of heterotrophs on readily degradable substrate and by growth of autotrophs. Therefore, to attain the same heterotrophic oxygen utilization as the conventional model, the IAWPRC Model has to process more readily degradable substrate through the synthesis/hydrolytic decay cycle. Because of this, the heterotrophic decay coefficient, b_H, and maximal specific growth rate constant, μ_m, applicable to the IAWPRC Model, are larger (by a factor of 2 to 3) than the corresponding values for the conventional model.

Table 2. Kinetics and Stiochiometry for the IAWPRC Model

Component →	1	2	3	4	5	6	7	8	9	10	11	12	13
1 Process	S_I	S_S	X_I	X_S	$X_{B,H}$	$X_{B,A}$	X_P	S_O	S_{NO}	S_{NH}	S_{ND}	X_{ND}	S_{ALK}
1 Aerobic growth of heterotrophs		$-\frac{1}{Y_H}$			1			$-\frac{1-Y_H}{Y_H}$		$-i_{XB}$			$\frac{-i_{XB}}{14}$
2 Anoxic growth of heterotrophs		$-\frac{1}{Y_H}$			1				$-\frac{1-Y_H}{2.86Y_H}$	$-i_{XB}$			$\frac{1-Y_H}{(14)\,2.86Y_H} - \frac{i_{XB}}{14}$
3 Aerobic growth of autotrophs						1		$-\frac{4.57-Y_A}{Y_A}$	$\frac{1}{Y_A}$	$i_{XB} - \frac{1}{Y_A}$			$-\frac{i_{XB}}{14} - \frac{1}{7Y_A}$
4 Decay of heterotrophs				$1-f_p$	-1		f_p					$i_{XB}-f_p i_{XP}$	
5 Decay of autotrophs				$1-f_p$		-1	f_p					$i_{XB}-f_p i_{XP}$	
6 Ammonification										1	-1		$\frac{1}{14}$
7 Hydrolysis		1		-1									
8 Hydrolysis of organic N											1	-1	

TABLE 3. RATE EXPRESSIONS FOR PROCESSES INCLUDED IN THE IAWPRC MODEL

	Process rate, $ML^{-3}T^{-1}$
1 Aerobic growth of heterotrophs	$\mu_{hm}\left(\frac{S_S}{K_S + S_S}\right)\left(\frac{S_O}{K_S+S_O}\right)X_{BH}$
2 Anoxic growth of heterotrophs	$\mu_{hm}\left(\frac{S_S}{K_S + S_S}\right)\left(\frac{K_{OH}}{K_{OH} + S_O}\right)\left(\frac{S_{NO}}{K_{NO} + S_O}\right)\eta_G X_{BH}$
3 Aerobic growth of autotrophs	$\mu_{am}\left(\frac{S_{NH}}{K_{NH} + S_{NH}}\right)\left(\frac{S_O}{K_{OA} + S_{NO}}\right)X_{BA}$
4 Decay of heterotrophs	$b_H X_{BH}$
5 Decay of autotrophs	$b_H X_{BH}$
6 Ammonification	$k_A S_{ND} X_{BH}$
7 Hydrolysis	$k_N \frac{X_S/X_{BH}}{K_X + (X_S/X_{BH})}\left[\left(\frac{S_O}{K_{OH} + S_O}\right) + \eta_N\left(\frac{K_{OH}}{K_{OH} + S_O}\right)\left(\frac{S_{NO}}{K_{NO} + S_{NO}}\right)\right]X_{BH}$
8 Hydrolysis of organic N	(process 7 rate) (X_{ND}/X_S)

Another difference between the IAWPRC and conventional models arises in the fractions of biomass decay converted to resistant particulate products. In the conventional model, this fraction is based on a "once through" decay and is generally taken as 0.2 (IAWPRC Task Group, 1986). However, the growth-decay processes are incorporated into a cycle in the IAWPRC Model, and f_p represents the fraction of decay converted to particulate products on each pass through the cycle. By considering the quantity of readily degradable substrate utilized and recycled through the growth/hydrolytic decay cycle, it can be reasoned that, for 1 gram of S_S used for growth,

$$Y_H(b/\mu)(1-f_p) \text{ grams of } S_S$$

are replenished through hydrolytic decay, giving a net substrate use of

$$1 - Y_H(b/\mu)(1-f_p)$$

TABLE 4. DEFINITION OF FOR THE IAWPRC MODEL

Component number	Component symbol	Definition
1	S_I	Soluble inert organic matter -- $M(COD)L^{-3}$
2	S_S	Readily biodegradable substrate -- $M(COD)L^{-3}$
3	X_I	Particulate inert organic matter -- $M(COD)L^{-3}$
4	X_S	Slowly biodegradable substrate -- $M(COD)L^{-3}$
5	X_{BH}	Active heterotrophic biomass -- $M(COD)L^{-3}$
6	X_{BA}	Active autotrophic biomass -- $M(COD)L^{-3}$
7	X_P	Particulate products arising from biomass decay -- $M(COD)L^{-3}$
8	S_O	Oxygen (negative COD) $M(COD)L^{-3}$
9	S_{NO}	Nitrate and nitrite nitrogen -- $M(N)L^{-3}$
10	S_{NH}	$NH_4^+ + NH_3$ nitrogen -- $M(N)L^{-3}$
11	S_{ND}	Soluble biodegradable organic nitrogen -- $M(N)L^{-3}$
12	X_{ND}	Particulate biodegradable organic nitrogen -- $M(N)L^{-3}$
13	S_{ALK}	Alkalinity--Molar units

Assuming that growth conditions are such that μ is approximately equal to b, gives,

$$\frac{b_C}{b} = \frac{f_p}{f_o} = (1 - Y_H (1 - f_p)) \qquad (15)$$

Taking the conventional f_o as 0.2, and the heterotrophic yield as 0.66, (based on biomass COD and substrate COD), gives $f_p = 0.078$, and 0.38 for the ratio of b_C/b.

Solution of the IAWPRC Model Using the SSSP Software Package

This software package was developed by Bidstrup and Grady (1987) and is available from Professor C.P.L. Grady at Clemson University, Clemson, South Carolina. A general description is given by Bidstrup and Grady (1988), and detailed instructions are given in the Users' Manual (Bidstrup and Grady,

TABLE 5. SUMMARY OF THE KINETIC AND STOICHIOMETRIC PARAMETERS COMTAINED IN THE IAWPRC MODEL

Kinetic parameters	Symbols
Heterotrophic growth and decay	μ_{hm} , K_S, K_{OH}, K_{NO}, b_H
Autotrophic growth and decay	μ_{am} , K_{NH}, K_{OA}, b_A
Correction factor for anoxic growth of heterotrophs	η_G
Ammonification	K_A
Hydrolysis	k_N, K_X
Correction factor for anoxic hydrolysis	η_H
Stoichiometric parameters	
Heterotrophic yield	Y_H
Autotrophic yield	Y_A
Fraction of biomass yielding particulate products	f_P
Mass N/Mass COD in biomass	i_{XB}
Mass N/Mass COD in products from biomass	i_{XP}

1987). It is written for the IBM Personal Computer or compatible machines and performs steady-state and dynamic simulations of activated sludge systems based on the IAWPRC Model. The program is user-friendly and versatile in that it allows the user to define the system configuration by using up to nine completely mixed reactors in series. Additional flexibility is provided by the capability to define the influent addition, return sludge recycle, and mixed liquor recirculation flow patterns between the various reactors.

The program contains "default" values of the stoichiometric and kinetic parameters required by the IAWPRC Model. These serve as reasonable starting points for simulations and calibration. One valuable feature of the program is that it allows the user to specify either reactor dissolved oxygen concentration or the reactor volumetric mass transfer coefficient, K_La. The mode in which the K_La value is specified is particularly useful for calibrations in which the reactor K_La is known. However, when K_La is

specified, the program uses identical values for the dissolved oxygen half-saturation constants for autotrophs and heterotrophs. This is a limitation because the value for autotrophs is generally believed to greater than the value for heterotrophs.

SECTION 5

TWENTY-FOUR HOUR PLANT STUDIES: DATA ACQUISITION AND MEASUREMENTS

ACTIVATED SLUDGE PLANTS USED FOR MODEL CALIBRATION

Six municipally owned and operated wastewater treatment plants were selected for calibration of the oxygen uptake model. All six of the plants were concurrently being studied as part of the ASCE-EPA Fine Bubble Diffused Aeration Design Manual Project (EPA, 1985) and, because of this, background information on process configuration and operation was readily available. Moreover, frequent off-gas measurements of oxygen transfer were being made at five of the six plants as part of the ASCE-EPA project. These measurements were particularly valuable because they produced accurate estimates of the process volumetric mass transfer coefficients for oxygen (K_La).

Twenty-four hour field studies were conducted at the plants described below. Additional information on the actual and modeled process flow diagrams and sampling locations is given in Section 7, Results and Model Calibration.

* Portage Lake Plant, Houghton, Michigan: This is an 8,700 m^3/d (2.3 MGD) contact stabilization activated sludge plant with aerobic sludge digestion and no primary sedimentation. It receives almost no industrial waste and was of value for model calibration because of the relatively low temperature and large particulate organic load in the influent. The plant was operated at a solids retention time of 10.6 days and produced a partially nitrified effluent. It was the only one of the six plants at which no off-gas measurements of oxygen transfer were available.

* Green Bay, Wisconsin: This is a modern 182,000 m^3/d (48 MGD) contact stabilization activated sludge plant receiving paper mill wastes which account for 30% of the flow and 50% of the 5 day BOD. The plant has a thermal sludge conditioning system and recycles the thermal sludge conditioning liquor to the activated sludge process. Other notable features of the plant include a high influent total Kjeldahl nitrogen concentration (40 to 60 mg/l) and warm process temperature. The plant operated at a solids retention time of 3.1 days and achieved very little nitrification. Off-gas measurements were made during the 24-hour study period.

* Madison, Wisconsin: This is a 151,000 m^3/d (40 MGD) activated sludge

plant receiving municipal wastes containing an appreciable (6% of flow, 15% of 5 day BOD) meat and cheese processing component. The plant operated at a solids retention time of 16.4 days and produced a nitrified effluent.

* Monroe, Wisconsin: This is an 8,330 m^3/d (2.2 MGD) step-feed activated sludge plant receiving a significant industrial load (17% of flow, 50% of 5 day BOD) consisting primarily of soluble cheese processing and brewing wastes. However, during the twenty-four hour study, an aerated in-line equalization basin was employed between the primary settling and activated sludge processes, and this significantly reduced the soluble COD fed to the activated sludge process. The plant operated at a solids retention time of 8.4 days and produced a nitrified effluent. Off-gas measurements were made during the 24-hour study period.

* Jones Island East Plant, Milwaukee, Wisconsin: This 288,000 m^3/d (76 MGD) plant receives an industrial load (11% of flow, 38% of 5 day BOD) dominated by the brewing, food processing and tanning industries. Primary treatment is by fine screening rather than sedimentation. The plant operated at a solids retention time of 2.8 days and experienced very little nitrification.

* South Shore Plant, Milwaukee, Wisconsin: This 371,000 m^3/d (98 MGD) plant receives a relatively light industrial load (6% of flow, 18% of 5 day BOD) from glue processing, food processing and machinery industries. Primary treatment is by sedimentation. The plant operated in a step feed configuration at a solids retention time of 4.3 days and achieved partial nitrification.

EXPERIMENTAL DESIGN OF 24-HOUR STUDIES

The 24-hour studies conducted at each of the plants were designed to acquire information necessary to calibrate the IAWPRC model. The steady-state model solution, based upon flow weighted average conditions, was employed for calibration of key parameters, and these parameters were subsequently used in the dynamic solution to simulate diurnal variations in oxygen uptake rate. Comparison of the simulated and measured diurnal ranges allowed assessment of the utility of the dynamic solution for design purposes. Accordingly, it was necessary to obtain plant operating data which would allow application of both the steady-state and dynamic solutions.

The dynamic solution contained in the SSSP software package requires information on process input concentrations and flows at intervals which are multiples of 15 minutes. It then uses linear interpolation to determine the values at the 15 minute sub-intervals. The sampling program was designed to develop the maximal amount of information within the limits of time available to the three person field study team. Key features of the sampling program included:

* activated sludge process influent and clarified effluent samples at two

hour intervals analyzed for: total COD, soluble COD, ammonia, organic nitrogen, nitrate, pH

* measurements at various positions in the aeration tank at hourly or longer intervals for: oxygen uptake rate (OUR), dissolved oxygen (DO), mixed liquor volatile suspended solids (MLVSS), temperature

In addition, the wastewater and recycle flow rates and the aeration rate were recorded at hourly intervals. The model predictions require that the process solids retention time (SRT) be established. This was determined based upon the sludge wasting practice in effect during the month preceding the 24-hour study. Normally, the SRT was fairly constant and an average of the daily values was used. However, at the Green Bay Plant, the solids wasting pattern varied considerably during the month preceding the study and the transient SRT was determined following the technique outlined by Baillod et al. (1977).

EXPERIMENTAL METHODS

Chemical Oxygen Demand (COD) was measured in the laboratory on preserved samples by the dichromate reflux method (Standard Methods, 1985).

Organic Nitrogen was measured in the laboratory on preserved samples by the macro-kjeldahl method (Standard Methods, 1985)

Ammonia Nitrogen was measured in the field using an Orion, Model 95-10, ammonia specific electrode.

Nitrate plus nitrite nitrogen was measured in the field following the recommended Hach procedure using Hach Nitraver V and a portable Hach DR-2 spectrophotometer.

Dissolved Oxygen (DO) was measured in the field using a Yellow Springs Instrument Company Model 54 dissolved oxygen probes. A submersible probe was employed to determine in-situ DO values in the aeration tanks.

Filtration for soluble COD and MLVSS measurement was performed using glass fiber filter papers as specified for suspended solids measurement (Standard Methods, 1985).

Oxygen Uptake Rate

Oxygen Uptake Rate was measured in the field using a batch BOD bottle technique.

a. A sample was withdrawn from the aeration tank using a weighted bucket and quickly transported to the field laboratory.

b. The sample was contacted with pure oxygen for a period of 5 to 20 seconds to raise the DO to about 10 mg/l.

c. An aliquot of the oxygenated mixed liquor was transferred to a 300 ml BOD bottle containing a magnetic stirring bar. A self-stirring YSI DO probe was inserted, and the bottle was submerged in a 2 liter container of mixed liquor to minimize temperature drift during the uptake measurement.

d. The two liter container was placed on a magnetic stirrer, and both the magnetic stirrer and self-stirring probe were turned on to provide mixing. The probe output of DO versus time was recorded using a Cole Parmer Model 8376-30 strip chart recorder. The oxygen uptake record was usually begun within two minutes of sample withdrawal from the aeration tank.

e. The OUR was determined from the slope of the strip chart record. The slope was determined from the earliest linear portion of the strip chart record. Normally this was during the first three minutes of record.

Interpretation of Oxygen Uptake Rates Measured by the BOD Bottle Method

The BOD Bottle Method described above has historically been used to measure oxygen uptake rates in activated sludge. The method itself gives an accurate and precise in-vitro measure of the oxygen utilization rate that occurs in the BOD bottle. The problem, however, is that this rate may not represent the in-situ oxygen utilization rate occurring in the reactor from which the sample was withdrawn. Two conditions, oxygen limitation and soluble substrate limitation, can cause the OUR measured by the BOD bottle method to differ significantly from the in-situ value.

Oxygen limitation arises when the in-situ DO is near zero and causes the in-situ oxygen uptake rate to be limited by the availability of oxygen. A sample subjected to the BOD bottle method is exposed to high DO concentrations and will respire at a higher rate. Consequently, for in-situ DO concentrations near zero, the OUR indicated by the BOD bottle method will be greater than the in-situ OUR. Substrate limitation arises when the in-situ exogenous substrate concentration is near zero and causes the OUR of a withdrawn sample to decrease between the time of withdrawal and the time at which the BOD bottle OUR is measured.

Mueller and Stensel (1987) compared in-situ OUR values estimated by non-steady state process water tests with those indicated by the bottle method at several activated sludge plants. The results showed that, at low DO values, the BOD bottle method produced indicated OUR values which tended to be higher than the in-situ values. At in-situ DO values below 1.5 mg/l, the indicated bottle uptake rates ranged from about 90% to 190% of the estimated in-situ OUR. The results also showed that soluble substrate depletion could produce indicated bottle OUR values significantly lower (by as much as 50%) than the estimated in-situ values. This was most prevalent in measurements made on completely mixed aeration tanks when the OUR was greater than about 250 mg/l/day. This was less prevalent in measurements made near the effluent end of long narrow aeration tanks.

In this study, the following strategy was used to cope with the inaccuracies inherent in using the BOD bottle method to estimate in-situ OUR

values:

* Off-gas measurements made at 5 of the 6 plants were used to estimate the process water volumetric mass transfer coefficients (K_La). Theoretically, this enabled the in-situ OUR to be determined from DO measurements and served as a check on the BOD bottle OUR. However, even though the off-gas measurements were made simultaneously with the 24-hour studies at two of the plants, changes in wastewater characteristics cause continual changes in the alpha factor and make it impossible to determine the process water K_La with a precision greater than 10%.

* BOD bottle OUR values measured on mixed liquor samples taken from regions where the DO was less than about 1.5 mg/l were not used for model calibration.

* More weight was given to BOD bottle OUR values measured on samples taken from zones near the effluent end of long narrow aeration tanks.

* Mixed liquor samples were withdrawn and handled rapidly so that OUR measurements were begun within 1 minute of sample withdrawal.

Off Gas Measurements

Off gas measurements were made by others at five of the six plants as part of a separate EPA Cooperative Agreement (EPA, 1985). These measurements followed the procedures of Redmon et al. (1983). Generally, the results of the off gas measurements were reported as Oxygen Transfer Efficiency (OTE) for various sections of the aeration tanks studied. These were converted to the process water volumetric mass transfer coefficient, K_La, values by,

$$K_La = \frac{OTE\ (W_{O2})}{V\ (C^*_\infty - S_O)} \qquad (16)$$

where:

W_{O2} = mass rate of oxygen applied to the tank section under consideration, M/t.

V = volume of the tank section under consideration, L^3

C^*_∞ = tank average oxygen saturation value approached at infinite time in the unsteady state clean water oxygen transfer test (ASCE, 1984)

S_O = average dissolved oxygen concentration in the tank section, M/L^3

SECTION 6

MODEL CALIBRATION

Calibration of the IAWPRC Model required that:

* the wastewater feed components required by the model be measured or estimated from the data collected during the plant studies,

* the kinetic and stoichiometric parameters required by the model be estimated,

* criteria for calibrating the model to the plant data be established.

The IAWPRC Task Group suggested methods for establishing values for the wastewater, stoichiometric, and kinetic parameters to be used in the model (IAWPRC Task Group, 1986)(Ekama et al., 1986). However, most of these methods are well suited only to laboratory pilot studies where the process is well controlled and significant effort can be spent on replicate measurements to assess and improve precision. The data in this report were based on 24-hour studies at six operating, full-scale municipal wastewater treatment plants. Although direct measurements of the wastewater feed components and model parameters following the recommendations of the Task Group were attempted, these techniques often were not practical for this study (e.g. time was not available for replicate measurements) or did not yield meaningful results (e.g. the anticipated step change in OUR required for S_S measurement did not occur). Because of these problems, modified methods were developed for measurement of feed components, and more reliance was placed on model calibration for parameter estimation.

DETERMINATION OF FEED COMPONENTS

The 13 components considered in the IAWPRC Model are defined in Table 4. Except for Component 7, particulate products arising from biomass decay, all of these components may be present in the feed wastewater. During the 24-hour plant studies, samples of the wastewater fed to and clarified effluent from the activated sludge processes were analyzed for total COD (TCOD), soluble COD (SCOD), ammonia, total organic nitrogen (TON), and nitrate. The following assumptions and logic were applied to the measured quantities to estimate the concentrations of the components required by the model.

Several components from Table 4 were assumed to be absent in the feed:

Component 5, X_{BH}, active heterotrophic biomass
Component 6, X_{BA}, active autotrophic biomass
Component 7, X_P, particulate products arising from biomass decay
Component 8, S_O, dissolved oxygen
Component 9, S_{NO}, nitrate plus nitrite

Measured values were used directly for Component 10, ammonia. Component 13, alkalinity, was taken as the annual average value for the plant. If the average annual value was not available, the default of 4 moles/m^3 was used. Lack of measured alkalinity values was not a serious drawback because the model calibration was not intended to include alkalinity. Moreover, the model does not include any relationships between alkalinity or pH and reaction rates.

The remaining six components, consisting of four COD fractions and two organic nitrogen fractions, were estimated from the measured data. It was assumed that the process removed essentially all of the biodegradable COD, so that the effluent soluble COD was taken as a measure of the inert soluble COD in the influent, $S_I = SCOD_{eff}$.

The inert particulate COD was assumed to account for between 10% and 20% of the influent particulate COD,

$$X_I = (0.1 \text{ to } 0.2)\ (TCOD_{inf} - SCOD_{inf}) \qquad (17)$$

The Task Group Report (IAWPRC, 1986) recommends that this component be estimated to fit solids production (MLVSS) data. Simulated values of the MLVSS concentration are sensitive to X_I, and higher percentages of the influent particulate COD (within the range of 10% to 20%) were selected as required to improve agreement between the simulated and measured concentrations.

The readily biodegradable COD, S_S, was estimated in two ways. First, it was assumed that it could be treated as if it were all soluble, thus,

$$S_S = SCOD_{inf} - S_I \qquad (18)$$

The second method (IAWPRC Task Group Report, 1986, Ekama et al., 1986) is based on the observation that, following a batch feeding of activated sludge with a wastewater containing both readily and slowly degradable substrates, the oxygen uptake rate (OUR) is maintained at a relatively high, constant level before abruptly declining to the lower endogenous level. The increase in OUR is assumed to be caused by the uptake of S_S at the rate U. Combining Equations 3 and 4 relates the increase in OUR to the substrate uptake rate in terms of the yield coefficient as,

$$U = \frac{\Delta r_O}{1 - Y_H} = \frac{\Delta \text{ OUR } / X_{BH}}{1 - Y_H} \qquad (19)$$

If the OUR is maintained at an elevated level for a time, t, before declining to the endogenous level, the readily biodegradable substrate, S_S, utilized, is given by

$$S_S = U X_{BH} t = \frac{\Delta \text{ OUR } t}{1 - Y_H} \quad (20)$$

Although the second method based on the batch reactor oxygen uptake data was applied to each of the six plants studied, only the Green Bay, Jones Island, and Portage Lake plants showed the anticipated step decrease in OUR and allowed estimation of S_S by this method. The other plants showed a gradual decrease in OUR and did not give meaningful results. Consequently, S_S was estimated by both methods for the Green Bay and Jones Island Plants and a separate model calibration was performed for each feed fractionation. Because both methods gave nearly identical (within 5%) S_S values for the Portage Lake Plant, only the first method, based on COD fractions, was used for the calibration. In all cases, the slowly degradable COD was determined by difference,

$$X_S = TCOD_{inf} - S_I - S_S - X_I \quad (21)$$

Similar logic was used to fractionate the organic nitrogen in the feed into four components (soluble biodegradable, S_{ND}, particulate biodegradable, X_{ND}, soluble inert, S_{NI}, and particulate inert, X_{NI}). It was assumed that the soluble effluent organic nitrogen consisted of inert substances. Thus, the inert soluble organic nitrogen in the influent, S_{NI}, was estimated by multiplying the total effluent organic nitrogen by the ratio of the effluent soluble COD to effluent total COD. The inert particulate organic nitrogen in the influent, X_{NI}, was estimated by multiplying the total organic nitrogen in the influent by the ratio of inert particulate COD in the influent to total COD in the influent. Thus, the influent biodegradable organic nitrogen was determined by subtracting the estimated values of the inert soluble and inert particulate organic nitrogen from the measured value of the total organic nitrogen. This was subdivided into soluble and particulate fractions by assuming that the organic nitrogen was divided in the same proportion as the COD.

$$S_{ND} = \frac{S_S}{S_S + X_S} (TON_{inf} - S_{NI} - X_{NI}) \quad (22)$$

The particulate biodegradable organic nitrogen in the influent was determined by difference,

$$X_{ND} = TON_{inf} - S_{NI} - X_{NI} - S_{ND} \quad (23)$$

DETERMINATION OF MODEL PARAMETERS

The effort here focused on the parameters to which the model simulations

are sensitive. To verify the predictive capability of a model, parameter estimates should be based on direct measurements. However, the IAWPRC model is based on fundamental stoichiometric, kinetic, and conservation principles that have been previously verified. Furthermore, it was pointed out earlier that, for these field studies, direct measurement was not practical. Consequently, sensitive parameters were determined by calibrating the model to the data. Parameters to which the model predictions were insensitive were determined from the literature and set at constant values.

This study focused mainly on the oxygen utilization aspects of the model. However, the model simulates many other system responses (e.g. mixed liquor volatile solids, dissolved oxygen, nitrate) in addition to oxygen uptake rate. Calibration of the model using only oxygen uptake rate data could be misleading. Consequently, an effort was made to calibrate the model so that it simulated MLVSS, dissolved oxygen, nitrate, and ammonia as well as oxygen uptake rate. Primary emphasis was placed on matching the oxygen uptake rate and DO. Lesser weight was placed on matching the MLVSS and nitrogen concentrations.

Sensitivity Analysis

Table 6 shows sensitivity coefficients for simulation of OUR, DO, MLVSS, ammonia, and nitrate to changes in key parameters. The parameters listed are the ones to which the simulations were most sensitive. This table was developed based on the simulation of the first reactor at the Monroe Plant for the default parameter estimates. The sensitivity coefficients were determined by changing each parameter by +20% from the default value while holding all other parameters fixed at their default levels. The dimensionless sensitivity coefficient is then expressed as

$$\text{Sensitivity Coefficient} = \frac{\text{percentage change in simulated value}}{\text{percentage change in key parameter}}$$

Positive sensitivity coefficients indicate an increase in the simulated value with an increase in the parameter value, whereas negative sensitivity coefficients indicate a decrease in the simulated value with an increase in the parameter value. Specific values of the sensitivity coefficients depend on features unique to the system being simulated. Table 6 pertains to reactor 1 of the Monroe Plant, and, because this plant had very high mixed liquor DO values, the simulated values were not sensitive to changes in either K_La or the half saturation constants for DO. Table 6 indicates that, of the biomass parameters, the OUR is most sensitive to the heterotrophic and autotrophic yield and decay coefficients. Overall, the heterotrophic yield, Y_H, and decay, b_h, showed the most impact on the simulated values of OUR, DO, and MLVSS. The autotrophic yield, Y_a, decay, b_a, and maximal specific growth rate, μ_{am}, had the most impact on simulated values of ammonia and nitrate. The last line in Table 6 shows the impact of the solids retention time (SRT), θ_c. Because SRT is an operational parameter and not a biomass parameter, it is shown separately. Values of SRT were not estimated as were the biomass parameters but were known from the operational data at each of the plants

TABLE 6. SENSITIVITY OF MODEL SIMULATIONS TO KEY PARAMETERS

Key Parameters	Sensitivity Coefficient for Simulation of				
	OUR	DO	MLVSS	S_{NH}	S_{NO}
μ_{hm}	+0.01	-0.1	0.0	0.0	-0.03
Y_H	-1.1	+0.7	+1.8	0.0	-0.5
b_h	+0.1	-0.8	-0.3	+0.4	-0.1
μ_{am}	+0.1	-0.1	0.0	-1.2	+0.1
Y_a	-0.2	+0.1	+0.04	+0.4	-0.6
b_a	-0.05	0.0	-0.01	+.8	-0.1
θ_c	+0.3	0.0	+0.6	-0.5	+0.15

studied. Table 6 indicates that the simulated values are sensitive to SRT, and, therefore, it is important that values of SRT be known accurately and precisely for the model simulations.

Parameter Estimation

The model parameters were estimated by calibrating the model to the average plant data measured during the 24-hour studies. The procedure was as follows:

1. Flow weighted average concentrations of the feed components required by the model were determined following the procedure described above.

2. The process flow diagram was modeled as realistically as possible using a combination of completely mixed reactors and perfect clarifier/thickeners. Tank volumes and flow rates were established based on plant data. Generally, preliminary decisions on reactor configurations for modeling were made in the field as data were collected. Further refinements took place during data analysis. The following criteria were collectively applied to divide the aeration tanks into completely mixed reactors for modeling:

   a) Variation of OUR and DO along the tank. When OUR and DO changed appreciably along the length of an aeration tank,

efforts were made to divide the tank into sections so that the actual values were within approximately 10% of the average value.

b) Points of feed addition. In cases where primary effluent was added at points along the aeration tank, it was convenient and logical to treat portions of the tank as separate reactors.

c) Aeration pattern. When tapered aeration was employed, it was convenient and logical to divide the tank into sections based on aeration intensity.

d) Length/width ratio. Long tanks were divided into individual reactors in an effort to keep the length/width ratio of the individual reactors at 5 or less.

3. A steady-state simulation was performed using the SSSP Software package with default values of all parameters. Table 7 summarizes the default values and approximate ranges (IAWPRC Task Group, 1986) for the parameters required by the model. A second steady-state simulation was performed using temperature adjusted default values for the kinetic parameters contained in the model. The expression,

$$\frac{\text{kinetic parameter at } T_2}{\text{kinetic parameter at } T_1} = 1.02^{(T_2-T_1)} \qquad (24)$$

was used to adjust all kinetic parameters to the average measured process temperature at each plant. The coefficient 1.02 is typical for the activated sludge process (Metcalf & Eddy, 1979).

4. Measured average values of reactor OUR, DO and MLVSS as well as effluent values of nitrate and ammonia were calculated for each of the plants.

5. The model simulated values of OUR, DO, MLVSS, nitrate and ammonia based on the temperature adjusted default parameters were compared with the measured average values. The model was then calibrated to the measured data by adjusting selected key model parameters. These adjustments were based on the sensitivity coefficients shown in Table 6, and the adjusted parameters were kept within the approximate range indicated in Table 7. Except for the Green Bay Plant, calibration adjustments were limited to the heterotrophic yield and decay parameters and the autotrophic maximal specific growth rate. In addition, the autotrophic half saturation constant for DO was made equal to the default value of the heterotrophic half saturation constant for DO (0.1 mg/l). The SSSP Software package requires that these two half saturation constants be equal when the simulation is based on given values of

TABLE 7. SUMMARY OF DEFAULT VALUES AND APPROXIMATE RANGES FOR MODEL PARAMETERS

Heterotrophic Parameter	Default Value	Approximate Range
Maximal Specific Growth Rate, μ_{hm}, d^{-1}	4.0	3 to 13
Half Saturation Constants		
Soluble Substrate, K_S, mg COD/l	10.	10 to 180
Dissolved Oxygen, K_{OH}, mg/l	0.1	0.1 to 0.15
Nitrate, K_{NO}, mg N/l	0.2	0.1 to 0.2
Yield Coefficient, Y_H, g COD/g COD	0.67	0.46 to 0.69
Decay Constant, b_H, d^{-1}	0.62	0.13 to 4.2
Anoxic Growth Factor, η_G,	0.8	0.6 to 1.0
Hydrolysis Rate, k_N, d^{-1}	2.2	-
Hydrolysis Saturation Ratio, K_X, g COD/g COD	0.15	-
Anoxic Hydrolysis Factor, η_H	0.4	-
Ammonification Rate, K_A, l/mg COD/d	0.16	-
Fraction of Biomass Yielding Particulate Products (X_P), f_p	0.08	-
Fraction N in Biomass, i_{XB}, g N/g COD	0.086	-
Fraction N in X_P, i_{XP}, g N/g COD	0.06	-
Autotrophic Parameter		
Maximal Specific Growth Rate, μ_{am}, d^{-1}	0.65	0.34 to 0.65
Half Saturation Constants		
Ammonia, K_{NH}, mg N/l	1.0	-
Dissolved Oxygen, K_{OA}, mg/l	1.0	0.1 to 2.0
Yield Coefficient, Y_A, mg COD/mg N	0.24	0.07 to 0.28
Decay Constant, b_A, d^{-1}	0.12	0.05 to 0.15

K_La in each reactor. The average DO saturation value for the tank was either determined from previous clean water oxygen transfer tests conducted at the plant or calculated from Equation D.2 of the ASCE Clean Water Standard (ASCE, 1984) with an effective saturation depth equal to one-third of the tank depth.

6. A dynamic simulation was performed using the SSSP Software package with the calibrated parameters determined from the steady state simulation. This gave simulated diurnal profiles for OUR and DO. These were then compared with the measured profiles to judge the utility of the model for simulating reasonable variations in oxygen utilization.

SECTION 7

RESULTS OF PLANT STUDIES AND MODEL SIMULATIONS

DATA DIRECTORY

The field study data and model calibration results for each of the six plants studied are presented and summarized in four types of tables. These are:

Plant Process Summary Tables: These contain information describing the wastewater characteristics and process configuration for each of the plants studied. (Tables 9, 11, 14, 16, 18, and 21)

Model Calibration Tables: These summarize and compare the average measured and steady-state simulation values of OUR, DO, ammonia, nitrate, and MLVSS along with the default, temperature adjusted default, and calibrated values of the model parameters. One model calibration table is given per plant, except for the Green Bay and Jones Island Plants where two different methods of feed COD fractionation resulted in two model calibration tables for each plant. (Tables 10, 12, 13, 15, 17, 19, 20, and 22)

Raw Data Tables: These contain the influent, effluent, and reactor data obtained for each plant during the 24-hour studies. They appear in Appendix A as Tables A1, A2, A3, A4, A5, and A6.

SSSP Steady State Simulation Tables: These summarize the essential input and output information for the steady-state simulations performed with the calibrated parameters. They appear in Appendix B as Tables B1, B2, B3, B4, B5, and B6.

SUMMARY OF MODEL CALIBRATION RESULTS

It was possible to obtain reasonable agreement between the average measured and steady-state simulated values of OUR, DO, MLVSS, nitrate, and ammonia at most of the plants studied. This agreement was achieved by calibrating only two out of three key parameters. The other parameters were set at their temperature adjusted default values. Notable lack of agreement between the simulated and measured values was evidenced at the Jones Island Plant where the simulated and measured effluent ammonia concentrations did not agree well, and at the South Shore Plant where the simulated and measured MLVSS values did not agree well.

Table 8 summarizes the values of the key parameters which resulted from model calibration. The key parameters were: the heterotrophic yield coefficient, Y_H, the heterotrophic decay constant, b_H, and the autotrophic maximal specific growth rate, μ_{am}.

The Green Bay plant differs from the others in that it is dominated by the heavy industrial load (pulp and paper). The soluble COD concentration in the feed to the activated sludge process at Green Bay was 718 mg/l. This was more than four times the average of the soluble COD concentration fed to the other five plants. Nevertheless, calibration was achieved by adjusting only two parameters when the feed fractions were based on COD data.

TABLE 8. SUMMARY OF MODEL CALIBRATION

Plant	All Parameters Were Set at Temperature Adjusted Default Values Except for:		
	Heterotrophic Yield Coefficient g COD/ g COD	Heterotrophic Decay Constant 1/d	Autotrophic Maximal Specific Growth Rate 1/d
Default (20°C)	0.67	0.62	0.65
Madison	0.57	d	0.34
Monroe	d	0.80	0.34
Portage Lake	0.69	1.10	d
Jones Island	d	0.35	0.62
South Shore	0.46	0.50	d
Green Bay	0.57	d	0.45

d signifies temperature adjusted default value

Interpretation of Steady State and Dynamic Simulations

In the following sections, steady state simulations are employed to calibrate the model parameters to the operating data collected at the six plants. The calibrated model is used to simulate dynamic behavior at three of the plants. In comparing the "calibrated" simulated values of DO, OUR, MLVSS, nitrate, and ammonia to the measured values perfect agreement should not be expected. Four reasons for this are: (1) the initial conditions upon which the simulation is based; (2) the fact that the plant surveys typically examined only one of several parallel aeration tanks contained in the total plant; (3) inaccuracies inherent in using the BOD bottle uptake

method to measure in-situ OUR; and (4) changes in the process water value of K_La.

To perform either a steady-state or dynamic simulation, the program must begin the simulation from a known starting or initial condition. The SSSP Software assumes that the flow and concentration values (in the case of the steady-state solution) or pattern (in the case of the dynamic solution) are constant from day to day. That is, the simulation assumes that the plant experienced conditions identical to the measured study conditions during the previous day, previous week, and previous month. In reality, inputs to municipal wastewater plants vary markedly with day of the week and season of the year. This will account for some lack of agreement between the measured and simulated values.

In the smaller plants studied, e.g. Portage Lake or Monroe, the aeration tanks studied accounted for one-third to one-half of the total biomass inventory and flow treated. However, in the large plants, such as Jones Island or South Shore, the aeration tanks studied accounted for only 4% to 5% of the biomass inventory and flow. Thus, the measurements in the smaller plants could be expected to be more representative of the average plant conditions.

In Section 5, it was pointed out that oxygen limitations arising when the in-situ DO value is near zero can cause the BOD bottle OUR method to yield erroneously high values (Mueller and Stensel, 1987). The simulations were based on values of the process water K_La measured by the off gas technique. For conditions where the simulated and measured DO values are equal, the simulated OUR value is derived entirely from the measured values of the process water K_La. It is subsequently shown that, for low DO conditions at the head of long aeration tanks, the SSSP Software correctly simulates in-situ OUR values much lower than the in-vitro values measured by the BOD bottle method.

It was also pointed out in Section 5 that soluble substrate limitations arising in completely mixed tanks could cause OUR values measured by the BOD bottle method to be somewhat lower than the true in-situ values. For these conditions, the model simulations were generally slightly higher than the OUR values measured by the bottle method.

At five of the six plants, K_La values based on off gas measurements (Redmon et al. 1983) were incorporated into the simulations. However, the process water K_La is not constant and may vary with time of day. Thus, even though the K_La measurements were sometimes made on the same day as the plant tests, the daily average process water K_La could have been different from the measured values. These differences would be reflected as differences between the measured and simulated OUR and DO values.

Portage Lake Plant

The plant process summary for the Portage Lake Plant is given in Table 9, the aeration tank configuration is shown in Figure 4, and the process flow diagram as modeled is shown in Figure 5. Table 10 shows the model

Table 9. PORTAGE LAKE PLANT PROCESS SUMMARY

Average Daily Flow: 8,700 m^3/d (2.3 MGD)
Average Raw Influent BOD_5: 150 mg/L
Major Industrial Contributors: none
Primary Treatment: coarse screening
Nominal SRT at Time of Study: 10.0 days
Recycle Ratio at Time of Study: 1.03
Fraction of Flow Treated by Aeration Basin Studied: 0.50

Process Configuration: This contact stabilization plant consists of 2 separate circular modular units, each of which includes a contact tank, reaeration basin, clarifier, and aerobic digestor. Both units were in operation during the study.

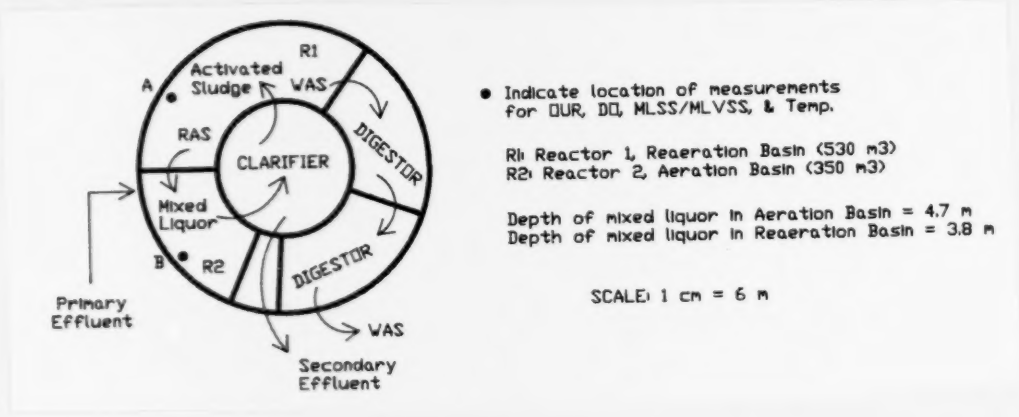

Figure 4. Portage Lake Plant plan sketch of basin one.

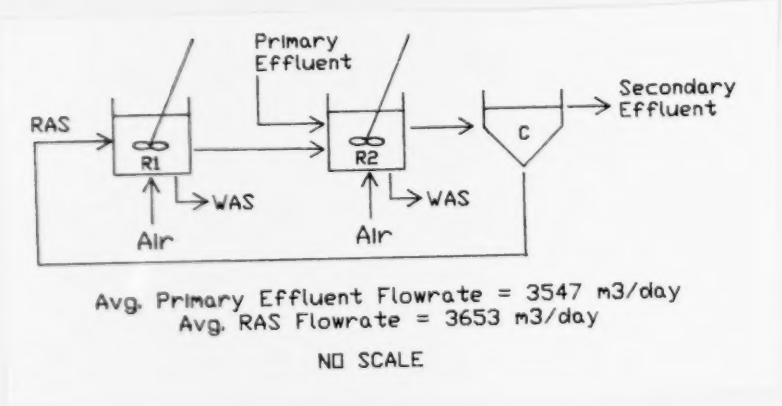

Figure 5. Portage Lake Plant process flow diagram as modeled.

calibration. For this plant, the default and temperature adjusted default values of the parameters produced remarkably good agreement between the measured and simulated values of OUR, DO, and MLVSS. The heterotrophic yield and decay coefficients were adjusted slightly to improve the agreement between the measured and simulated nitrate concentrations. The simulated values of the OUR tend to be greater than the measured values. However, for the well mixed contact tank, substrate depletion may have caused the BOD bottle method to underestimate the in-situ OUR.

Figure 6 compares the measured and dynamically simulated values of OUR and DO for the Portage Lake contact tank. The agreement between the OUR values is surprisingly good. However, the DO comparison suggests that the model may simulate changes before they actually occur. Nevertheless, the model did simulate the range and trend of OUR values and the range of DO values.

Green Bay Plant

Table 11 gives the Green Bay Plant Process Summary. Figure 7 shows the aeration tank configuration and Figure 8 shows the process flow diagram as modeled. A notable feature is that 50% of the BOD results from paper mill wastes. The model calibration shown in Table 12 indicates that the default and temperature adjusted default simulations gave values of OUR close to the measured values, and values of MLVSS somewhat higher than the measured values. Values of ammonia were considerably out of agreement with the measured values. In addition, the simulation results showed considerable denitrification. The simulated MLVSS values were reduced by decreasing the heterotrophic yield, and the simulated ammonia values were increased by decreasing the autotrophic maximal specific growth rate.

The average of the daily SRT values during the previous month was 8.3 days. However, this was influenced by a week of very high (13d to 62d) daily SRT values which occurred early in the preceding month. During the ten days preceding the field study, the daily SRT values ranged from 2.4 to 3.1 days. The actual transient SRT at the time of the field study was estimated as 3.1 days following the method given by Baillod et al. (1977).

Green Bay was one of the two plants in which the method based on the increase and subsequent decrease in OUR observed in a batch reactor following feeding gave meaningful results for estimation of readily degradable substrate (IAWPRC Task Group, 1986). Table 13 shows the model calibration for the feed COD fractionation based on the batch reactor data. Although the same values of the model parameters were used for both Tables 12 and 13, differences in the feed fractions (Table 12 is based on S_S = 510, and X_S = 278, whereas Table 13 is based on S_S = 253 and X_S = 536.) caused slight differences in the simulated values.

In both Tables 12 and 13, the simulated values of the OUR are considerably greater than the measured values. However, these measurements are in a large well-mixed tank, a condition that would be conducive to substrate depletion which would cause the BOD bottle method to give low measured values for OUR. Moreover, the good general agreement between the

TABLE 10. MODEL CALIBRATION FOR THE PORTAGE LAKE PLANT

SSSP STEADY-STATE SIMULATION

SRT = 10.6 days Temperature = 18.0 C		Measured Values	SIMULATED VALUES Default	Default w/ Temp. Adj.	Calibrated
OUR (mg O2/L/day):Contact Tank		598	600.6	663.6	665.2
DO (mg/L):	Contact Tank	0.5	0.6	2.0	1.2
	Reaeration Tank	1.5	0.6	1.6	1.5
Effluent Ammonia (mg-N/L)		2.6	1.9	0.9	0.7
Effluent NO3/NO2 (mg-N/L)		4.5	0.0	0.0	1.7
MLVSS (mg COD/L): Contact Tank		2361	2438.6	2344.9	2156.7
K(L)a (1/day):	Contact Tank	72	72	72	72
	Reaeration Tank	72	72	72	72

VALUES OF MODEL PARAMETERS USED FOR SIMULATIONS

	μ max	1/d	4.000	3.845	3.845
H	Ks COD	mg COD/L	10.0	9.612	9.612
E	Ks O2	mg O2/L	0.55		0.10
T	yield	mg cells/mg COD	0.67		0.69
E	b decay	1/d	0.620	0.596	1.100
R	anoxic growth factor		0.8		
O	Ks NO3	mg-N/L	0.2		
T	hydrolysis rate	1/d	2.200	2.115	2.115
R	hydrol satur ratio	g COD/g COD	0.150	0.156	0.156
O	anoxic hydrol factor		0.4		
P	ammonification	L/mg COD/d	0.160	0.154	0.154
H	frac. part. prod.	mg COD/mg COD	0.080		
I	N in biomass	mg-N/mg COD	0.086		
C	N in part. prod.	mg-N/mg COD	0.060		
	O2 saturation conc	mg-O2/L	9.0	11.0	11.0
A	μ max	1/d	0.650	0.625	0.625
U	Ks NH-N	mg-N/L	1.000		
T	Ks O2	mg-O2/L	0.550		0.10
O	yield	mg cells/mg COD	0.240		
	b decay	1/d	0.120		

measured and simulated DO values (i.e. the measured and simulated DO driving force for mass transfer) means that the OUR simulation is in agreement with the off-gas measurements upon which K_La is based.

Madison Plant

The Madison Plant Process Summary is shown in Table 14. Figure 9 shows the configuration of the basins studied and Figure 10 shows the process flow diagram as modeled. A distinguishing feature of this plant is the long SRT (16.4 days) and resulting high degree of nitrification. The plant was modeled as four reactors in series, and only reactors 3 and 4, where the average DO values were above 3 mg/l, were used to calibrate the model. Table 15 indicates that the temperature adjusted default parameter values gave high

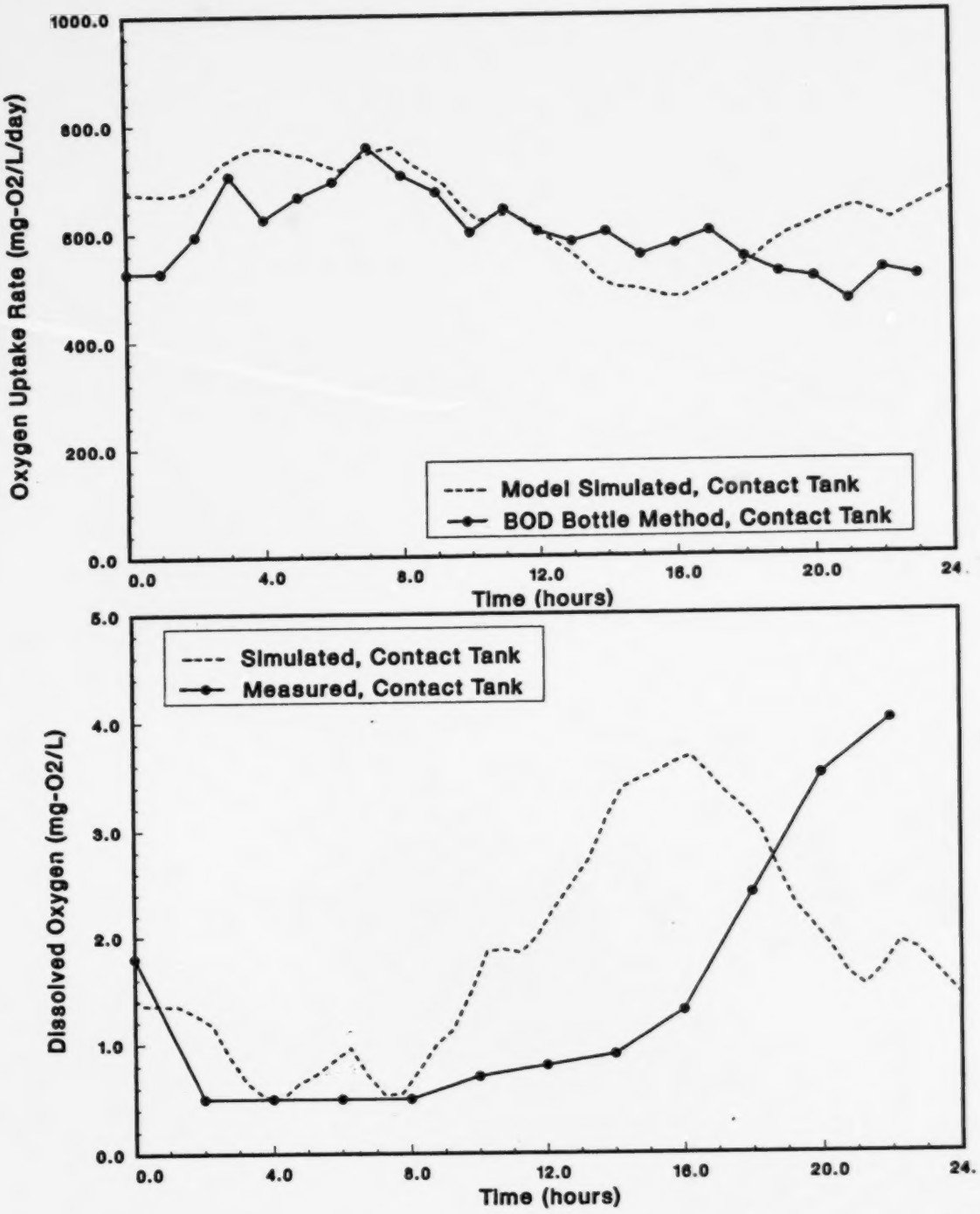

Figure 6. Portage Lake Plant dynamic simulation: comparison of measured and simulated OUR and DO values.

Table 11. GREEN BAY PLANT PROCESS SUMMARY

Average Daily Flow: 182,000 m^3/d (48 MGD)
Average Raw Influent BOD_5: 375 mg/L
Major Industrial Contributors: paper mills
Fraction of Flow: 0.30
Fraction of BOD_5: 0.50
Primary Treatment: sedimentation
Nominal SRT at Time of Study: 3.1 days
Recycle Ratio at Time of Study: 0.66
Fraction of Flow Treated by Aeration Basin Studied: 0.33

Process Configuration: This plant consists of 4 parallel contact stabilization processes, each of which includes an aeration (contact) basin, approximately 74 m. (245 ft.) long and 22 m. (74 ft.) wide. Three of these processes were in operation during the study.

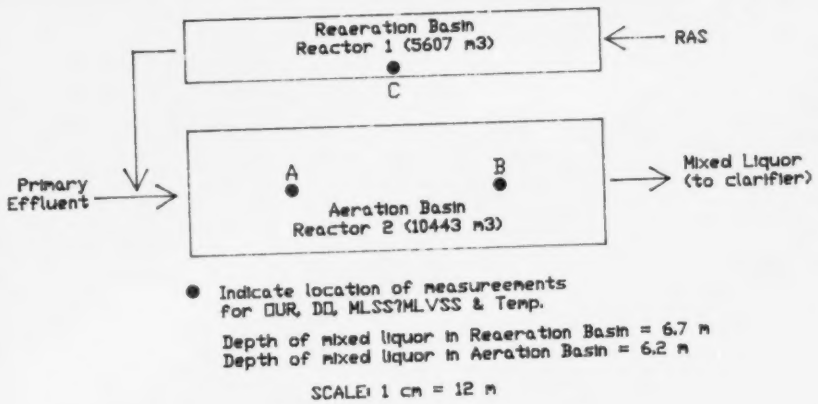

Figure 7. Green Bay Plant plan sketch of basin four.

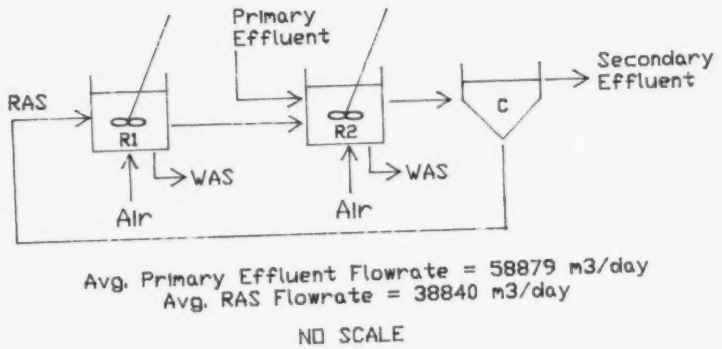

Figure 8. Green Bay Plant process flow diagram as modeled.

TABLE 12. MODEL CALIBRATION FOR THE GREEN BAY PLANT FEED FRACTION BASED ON COD FRACTIONS

SRT = 3.1 days Temperature = 26.5 C		Measured Values	SIMULATED VALUES Default	Default w/ Temp. Adj.	Calibrated
OUR (mg O2/L/day) Contact Tank		1840	1752	1879	2201
DO (mg/L):	Reaeration Tank	0.3	0.4	0.4	0.1
	Contact Tank	1.5	3.3	3.6	2.5
Effluent Ammonia (mg-N/L)		30.4	19.6	13.5	32.2
Effluent NO3/NO2 (mg-N/L)			0	0.2	0
MLVSS (mg COD/L): Contact Tank		2792	3197	3066	2446
K(L)a (1/day):	Reaeration Tank	118	118	118	118
	Contact Tank	311	311	311	311

VALUES OF MODEL PARAMETERS USED FOR SIMULATIONS					
H	μ max	1/d	4.000	4.549	4.549
E	Ks cod	mg COD/L	10.0	11.374	11.374
T	Ks O2	mg-O2/L	0.55		0.1
E	yield	mg cells/mg COD	0.67		0.57
R	b decay	1/d	0.620	0.705	0.705
O	anoxic growth factor		0.8		
T	Ks NO3	mg-N/L	0.2		
R	hydrolysis rate	1/d	2.200	2.502	2.502
O	hydrol satur ratio	g COD/g COD	0.150	0.132	0.132
P	anoxic hydrol factor		0.4		
H	ammonification	L/mg COD/d	0.160	0.182	0.182
I	frac. part. prod.	mg COD/mg COD	0.080		
C	N in biomass	mg-N/mg COD	0.086		
	N in part. prod.	mg-N/mg COD	0.060		
	O2 saturation conc	mg O2/L	9.0	9.7	9.7
	μ max	1/d	0.650	0.739	0.45
A	Ks NH-N	mg-N/L	1.000		
U	Ks O2	mg-O2/L	0.550		0.1
T	yield	mg cells/mg COD	0.240		
O	b decay	1/d	0.120		

simulated values for DO and MLVSS, and low simulated values for OUR. Calibration required that the yield coefficient be decreased to increase the OUR and decrease the MLVSS. Simultaneously, the maximal autotrophic specific growth rate was decreased. The simulated DO values agreed fairly well (within 0.5 mg/l) with the measured values, whereas the simulated OUR values for Reactors 3 and 4 were 5% to 10% below the measured values.

Figure 11 shows the dynamic simulation of the OUR and DO values along with the measured values for Reactors 3 and 4. Except for the last four hours of the study, the measured and simulated values agree remarkably well. Moreover, the model accurately simulated the range of diurnal DO change in Reactors 3 and 4. The simulated and measured profiles of OUR and DO along the tank length are given in Figure 12. The OUR profile clearly shows the simulated values in Reactors 1 and 2 to be well below the corresponding values measured by the BOD bottle method. The general agreement between the simulated and

TABLE 13. MODEL CALIBRATION FOR GREEN BAY PLANT WITH FEED FRACTIONS BASED ON BATCH REACTOR DATA

SRT = 3.1 days Temperature = 26.5 C		Measured Values	SIMULATED VALUES Default	 Default w/ Temp. Adj.	 Calibrated
OUR (mg O2/L/day) Contact Tank		1840	1452	1599	2201
DO (mg/L):	Reaeration Tank	0.3	0.3	0.3	0
	Contact Tank	1.5	4.2	4.4	2.5
Effluent Ammonia (mg-N/L)		30.4	21.4	19.1	32.7
Effluent NO3/NO2 (mg-N/L)			0	0.1	0
MLVSS (mg COD/L): Contact Tank		2792	3578	3365	2439
K(L)a (1/day):	Reaeration Tank	118	118	118	118
	Contact Tank	311	311	311	311

VALUES OF MODEL PARAMETERS USED FOR SIMULATIONS					
H	μ max	1/d	4.000	4.549	4.549
E	Ks cod	mg COD/L	10.0	11.374	11.374
T	Ks O2	mg-O2/L	0.55		0.1
E	yield	mg cells/mg COD	0.67		0.57
R	b decay	1/d	0.620	0.705	0.705
O	anoxic growth factor		0.8		
T	Ks NO3	mg-N/L	0.2		
R	hydrolysis rate	1/d	2.200	2.502	4.110
O	hydrol satur ratio	g COD/g COD	0.150	0.132	0.132
P	anoxic hydrol factor		0.4		
H	ammonification	L/mg COD/d	0.160	0.182	0.182
I	frac. part. prod.	mg COD/mg COD	0.080		
C	N in biomass	mg-N/mg COD	0.086		
	N in part. prod.	mg-N/mg COD	0.060		
	O2 saturation conc	mg O2/L	9.0	9.7	9.7
	μ max	1/d	0.650	0.739	0.45
A	Ks NH-N	mg-N/L	1.000		
U	Ks O2	mg-O2/L	0.550		0.1
T	yield	mg cells/mg COD	0.240		
O	b decay	1/d	0.120		

measured DO values in Reactors 1 and 2 indicates that the OUR is being correctly simulated based on the measured K_La, and that the apparent discrepancy between the simulated and measured OUR values in Reactors 1 and 2 is an artifact of in-situ oxygen limitations on the BOD bottle uptake rate measurement method.

Monroe Plant

The Process Summary for the Monroe Plant is given in Table 16. Figure 13 shows the configuration of the aeration basin studied and Figure 14 shows the process flow diagram as modeled. Distinguishing features of this plant are the dominant industrial contributions of the brewing, dairy, and food processing industries and the in-line, aerated flow equalization basin. This basin functioned to reduce the high soluble COD and to increase the particulate COD in the wastewater fed to the activated sludge process. Another

Table 14. MADISON PLANT PROCESS SUMMARY

Average Daily Flow (East and West Plants): 151,000 m^3/d (40 MGD)
Average Raw Influent BOD_5: 170 mg/L
Major Industrial Contributors: meat and cheese processors
Fraction of Flow: 0.06
Fraction of BOD_5: 0.15
Primary Treatment: sedimentation
Nominal SRT at Time of Study: 16.4 days
Recycle Ratio at Time of Study: 0.67
Fraction of West Plant Flow* Treated by Aeration Basin Studied: 0.45
* Average flow for the West Plant during the study was 45,370 m^3/d (11.99 MGD).

Process Configuration: The plant is divided into two sub plants, East and West. The West Plant, in turn is divided into Units 3 and 4. This study was conducted on Unit 3 which includes 2 three-pass aeration tanks, with each pass 80.8 m. (265 ft.) long and 9.2 m. (30 ft.) wide. Effluent from both tanks of Unit 3 is combined before clarification.

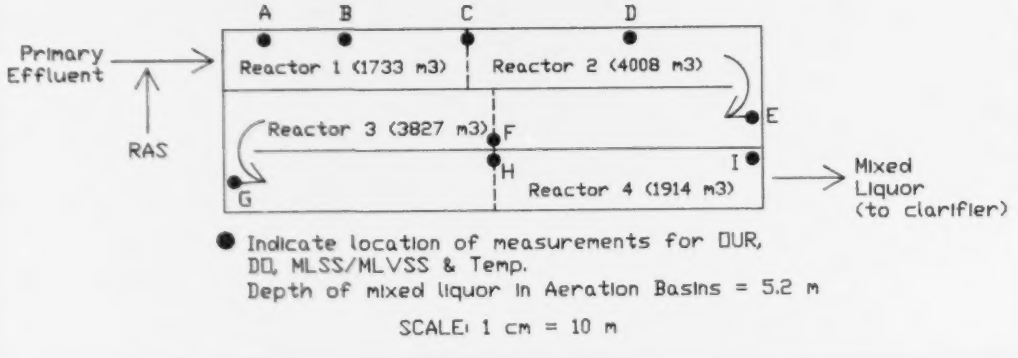

Figure 9. Madison Plant plan sketch of basins 22, 23 and 24.

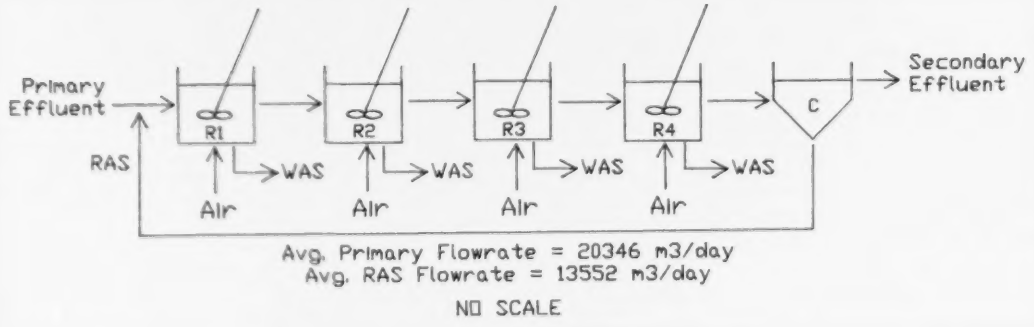

Figure 10. Madison Plant process flow diagram as modeled.

TABLE 15. MODEL CALIBRATION FOR THE MADISON PLANT

SRT = 16.4 days Temperature = 20.5 C		Measured Values	SIMULATED VALUES Default	Default w/ Temp. adj.	Calibrated
OUR (mg O2/L/day)	Reactor 3	447	337.4	330.7	404.8
	Reactor 4	363	302.4	295.4	341.3
DO (mg/L)	Reactor 3	3.0	2.1	3.7	2.4
	Reactor 4	3.9	2.8	4.5	3.4
Effluent Ammonia (mg-N/L)		0.2	0.2	0.2	0.6
Effluent NO3/NO2 (mg-N/L)		12	6.3	8.1	11.7
MLVSS (mg COD/L)	Reactor 1	2194	2280.1	2263.5	1897.8
K(L)a (1/day):	Reactor 1	45-55*	45	45	45
	Reactor 2	68-77*	68	68	68
	Reactor 3	43-50*	50	50	50
	Reactor 4	37-51*	51	51	51

VALUES OF MODEL PARAMETERS USED FOR SIMULATIONS					
H	μ max	1/d	4.00	4.04	4.04
E	Ks COD	mg COD/L	10.0	10.1	10.1
T	Ks O2	mg O2/L	0.55		0.10
E	yield	mg cells/mg	0.67		0.60
R	b decay	1/d	0.620	0.626	0.626
O	anoxic growth factor		0.8		
T	Ks NO3	mg-N/L	0.2		
R	hydrolysis rate	1/d	2.20	2.222	2.222
O	hydrol satur ratio	g COD/g COD	0.150	0.149	0.149
P	anoxic hydrol factor		0.4		
H	ammonification	L/mg COD/d	0.160	0.162	0.162
I	frac. part. prod.	mg COD/mg CO	0.080		
C	N in biomass	mg-N/mg COD	0.086		
	N in part. prod.	mg-N/mg COD	0.060		
	O2 saturation conc	mg O2/L	9.0	10.5	10.5
	μ max	1/d	0.650	0.656	0.34
A	Ks NH-N	mg-N/L	1.000		
U	Ks O2	mg O2/L	0.550		0.10
T	yield	mg cells/mg	0.240		
O	b decay	1/d	0.120		

notable feature is the high DO levels (greater than 6 mg/l) maintained in the shown in Table 17 gave simulated values of OUR which were somewhat high in zone 1 and somewhat low in zone 2. Agreement between the simulated and measured values was improved by slightly increasing the heterotrophic decay rate and decreasing the autotrophic maximal specific growth rate.

The dynamic simulation given in Figure 15 shows good general agreement between the measured and simulated values. Moreover, the in-line equalization basin effectively attenuated the diurnal variations.

Jones Island Plant

Table 18 shows the process summary for the Milwaukee Jones Island East Plant. Figure 16 shows the configuration of the tank studied and Figure 17

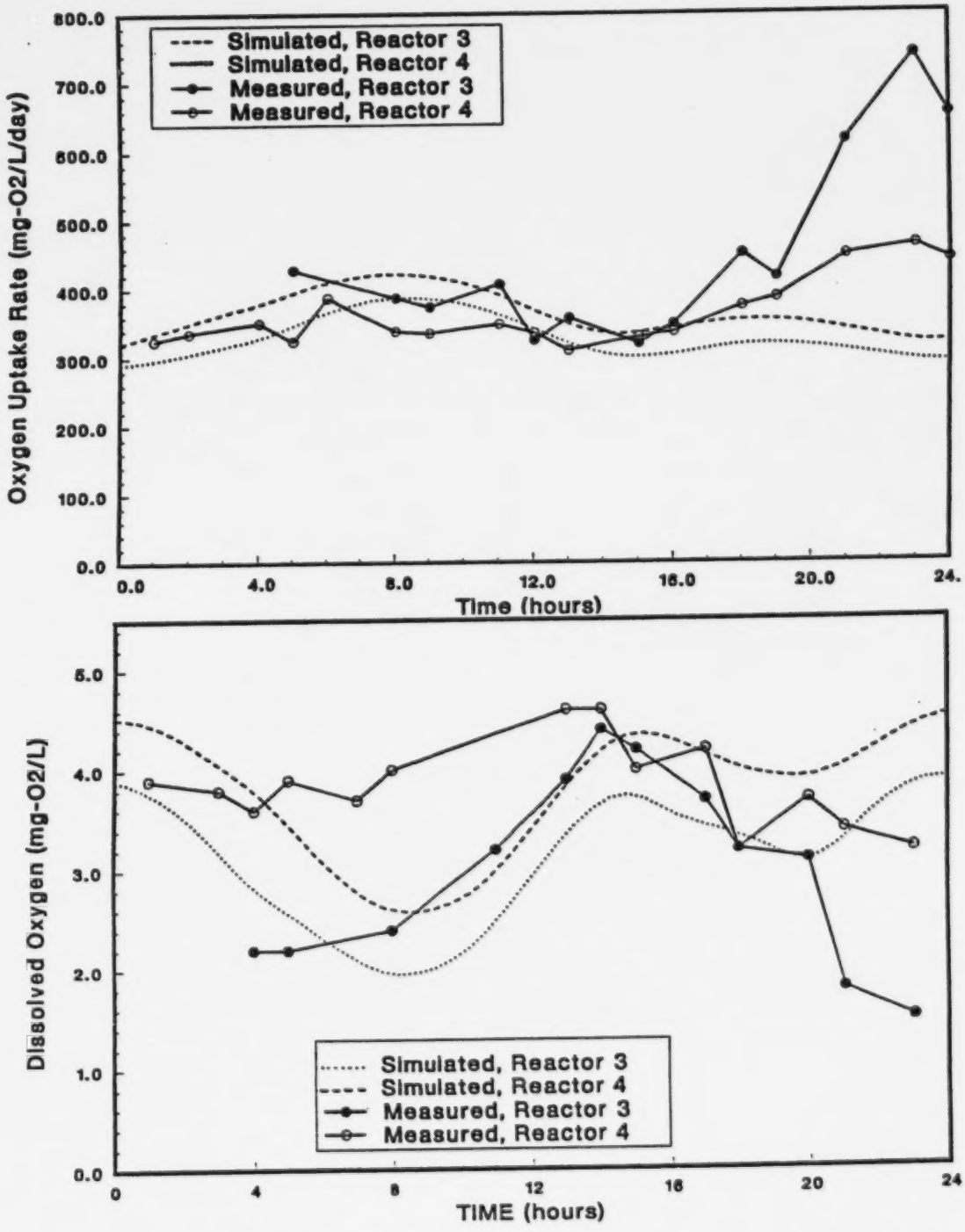

Figure 11. Madison Plant dynamic simulation: comparison of measured and simulated OUR and DO values.

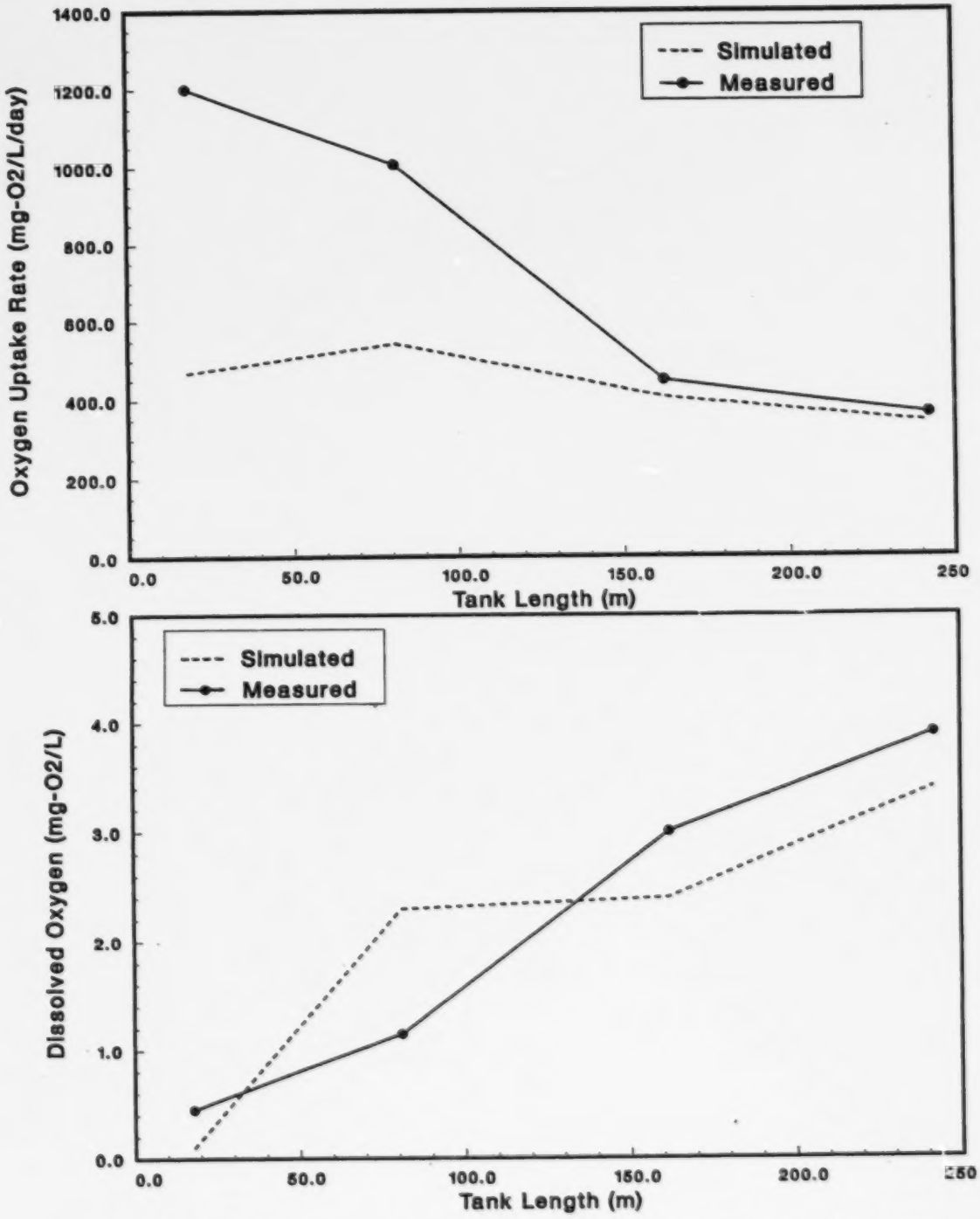

Figure 12. Madison Plant: variation of average OUR and DO values along tank length.

Table 16. MONROE PLANT PROCESS SUMMARY

Average Daily Flow: 8,330 m^3/d (2.2 MGD)
Average Raw Influent BOD_5: 418 mg/L
Major Industrial Contributors: breweries and dairy and food processors
Fraction of Flow: 0.17
Fraction of BOD_5: 0.50
Primary Treatment: sedimentation with flow equalization
Nominal SRT at Time of Study: 8.4 days
Recycle Ratio at Time of Study: 1.25
Fraction of Flow Treated by Aeration Basin Studied: 0.33

Process Configuration: This plant includes 3 two-pass basins, with each pass 31.1 m. (102 ft.) long and 7.6 m. (25 ft.) wide. The basins are fed via step feed in the first pass. Mixed liquor from the aeration basins is combined before clarification. All three basins were in operation during the study.

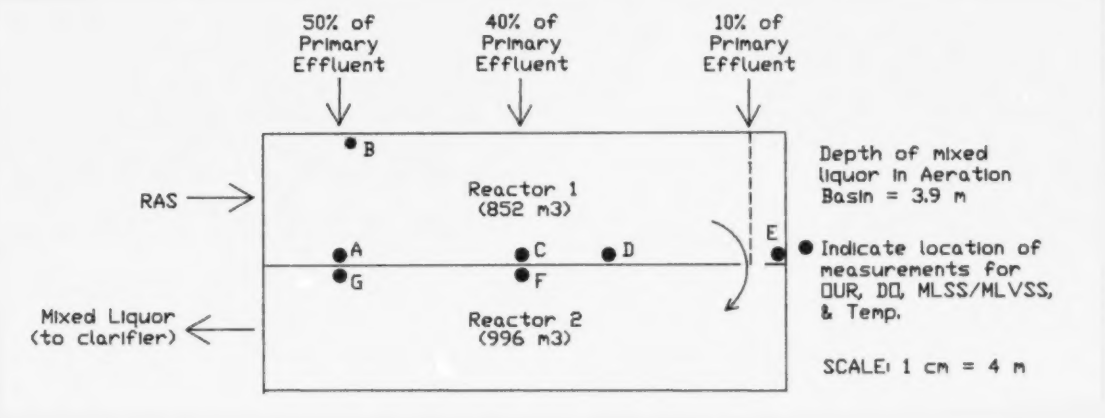

Figure 13. Monroe Plant plan sketch of basin three.

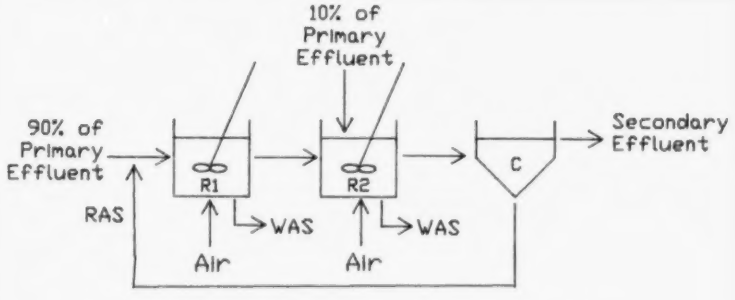

Figure 14. Monroe Plant process flow diagram as modeled.

TABLE 17. MODEL CALIBRATION FOR THE MONROE PLANT

SRT = 8.4 days Temperature = 23.1 C		Measured Values	SIMULATED VALUES Default	Default w/ Temp. adj.	Calibrated
OUR (mg O2/L/day)	Zone 1	478	412.8	523.2	497.9
	Zone 2	418	292.9	352.7	398.8
DO (mg/L):	Zone 1	6.1	6	6.1	6.3
	Zone 2	6.5	6.4	6.7	6.3
Effluent Ammonia (mg-N/L)		0.1	0.4	0.4	2
Effluent NO3/NO2 (mg-N/L)		11.5	14.9	17.1	19.1
MLVSS (mg COD/L):	Zone 1	1822	2229.9	1686.7	1613.9
	Zone 2	1491	2112.4	1587.7	1516.8
K(L)a (1/day):	Zone 1	155	155	155	155
	Zone 2	115	115	115	115
VALUES OF MODEL PARAMETERS USED FOR SIMULATIONS					
H	μ max	1/d	4.000	4.253	4.253
E	Ks COD	mg COD/L	10.0	10.633	10.633
T	Ks O2	mg-O2/L	0.55		0.10
E	yield	mg cells/mg COD	0.67		0.67
R	b decay	1/d	0.620	0.659	0.800
O	anoxic growth factor		0.8		
T	Ks NO3	mg-N/L	0.2		
R	hydrolysis rate	1/d	2.200	2.339	2.339
O	hydrol satur ratio	g COD/g COD	0.150	0.141	0.141
P	anoxic hydrol factor		0.4		
H	ammonification	L/mg COD/d	0.160	0.170	0.170
I	frac. part. prod.	mg COD/mg COD	0.080		
C	N in biomass	mg-N/mg COD	0.086		
	N in part. prod.	mg-N/mg COD	0.060		
	O2 saturation conc	mg-O2L	9.00	9.82	9.82
	μ max	1/d	0.650	0.691	0.340
A	Ks NH-N	mg-N/L	1.000		
U	Ks O2	mg-O2/L	0.550		0.100
T	yield	mg cells/mg COD	0.240		
O	b decay	1/d	0.120		

shows the process flow diagram as modeled. Notable features of this plant are the relatively high (38% of BOD) industrial waste component, high average influent concentration (300 mg/l 5 day BOD), and short SRT (2.8 days). The long, single pass tank studied was modeled as a series of 5 completely mixed reactors in series.

Tables 19 and 20 show the model calibrations for feed COD fractions based on measured soluble and particulate components and batch reactor data respectively. (For Table 19, S_S = 337, and X_S = 139; for Table 20, S_S = 151, and X_S = 325.) Oxygen uptake rate (OUR) and DO data from reactors two and four were used for model calibration. It was apparent that the extremely high bottle OUR values measured for reactor one were not representative of the in-situ values at the prevailing low DO conditions in that reactor. The values of OUR, DO, and MLVSS simulated by the default parameters for reactors 2 and 4 agreed reasonably well with the measured values, and this agreement was

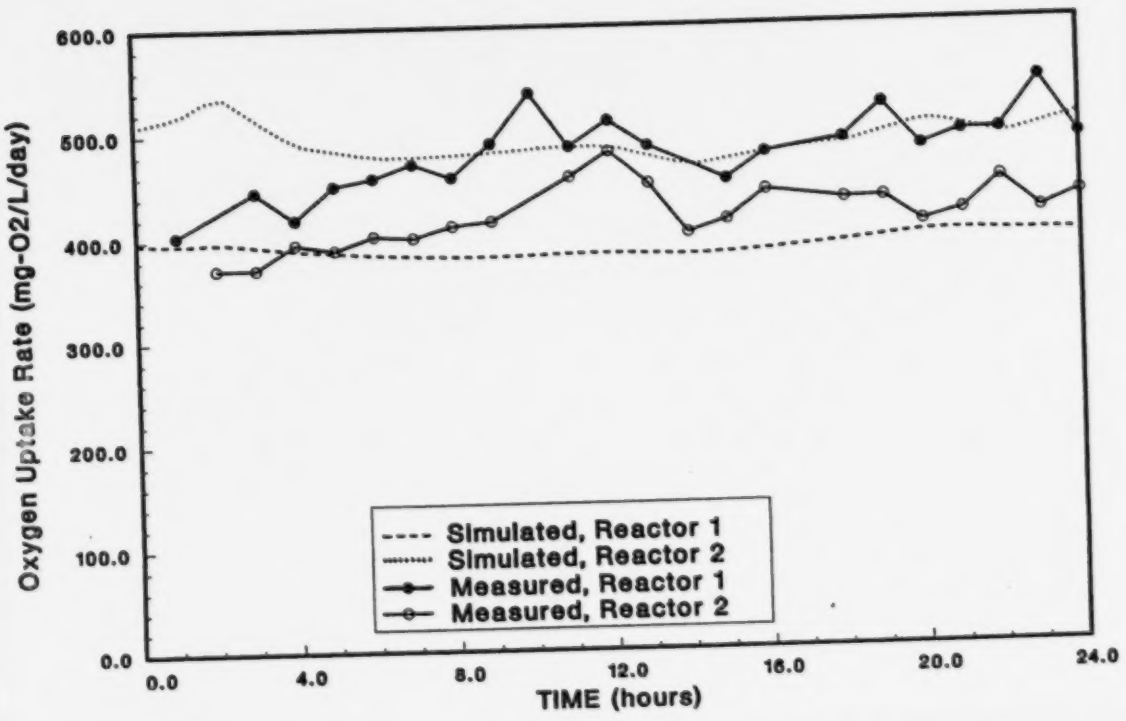

Figure 15. Monroe plant dynamic simulation: comparison of measured and simulated OUR values.

improved by adjusting the heterotrophic decay coefficient and maximal autotrophic specific growth rate.

However, the measured and simulated ammonia concentrations did not agree. This plant had relatively high concentrations of ammonia and organic nitrogen in the feed (16.7 and 30.6 mg/l respectively) but the effluent ammonia concentration was measured to be only 5 mg/l. At the operating SRT of 2.8 days, the model simulated effluent ammonia concentrations ranged from 12 to 21 mg/l. The simulation with feed fractions based on the batch reactor data (Table 20) simulated more ammonia loss through nitrification-denitrification and consequently produced better agreement between the measured (5 mg/l) and simulated (12 mg/l) ammonia values.

Figure 18 shows the diurnal profile for observed and measured values of OUR and DO. The observed and simulated values of DO agree reasonably well, and this means that the simulated OUR values should be realistic. (Recall that, since the simulation is based on K_La measured by off-gas, perfect agreement between the measured and simulated DO values means that the OUR

Table 18. JONES ISLAND EAST PLANT PROCESS SUMMARY

Average Daily Flow: 288,000 m^3/d (76 MGD)
Average Raw Influent BOD_5: 300 mg/L
Major Industrial Contributors: breweries, food processors, tanneries
Fraction of Flow: 0.11
Fraction of BOD_5: 0.38
Primary Treatment: fine screening
Nominal SRT at Time of Study: 2.8 days
Recycle Ratio at Time of Study: 0.33
Fraction of Flow Treated by Aeration Basin Studied: 0.04

Process Configuration: The East Plant includes 20 aeration tanks each divided into 2 parallel one-pass basins, with each pass 111 m. (360 ft.) long and 7.5 m. (24.5 ft.) wide. Effluent from a group of basins flows to a cluster of common clarifiers. There were 14 tanks (28 basins) in operation during the 24-hour study, and the north basin of Tank # 6 was monitored.

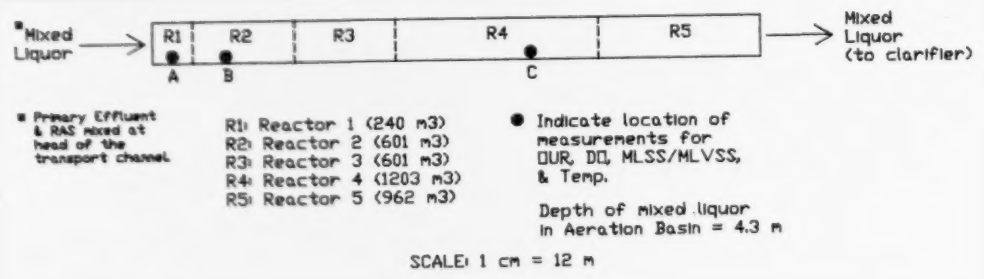
Mixed Liquor
R1
R2
R3
R4
R5
Mixed Liquor (to clarifier)
A
B
C
Primary Effluent & RAS mixed at head of the transport channel
R1: Reactor 1 (240 m3)
R2: Reactor 2 (601 m3)
R3: Reactor 3 (601 m3)
R4: Reactor 4 (1203 m3)
R5: Reactor 5 (962 m3)
Indicate location of measurements for OUR, DO, MLSS/MLVSS, & Temp.
Depth of mixed liquor in Aeration Basin = 4.3 m
SCALE: 1 cm = 12 m

Figure 16. Jones Island Plant plan sketch of north basin of tank six.

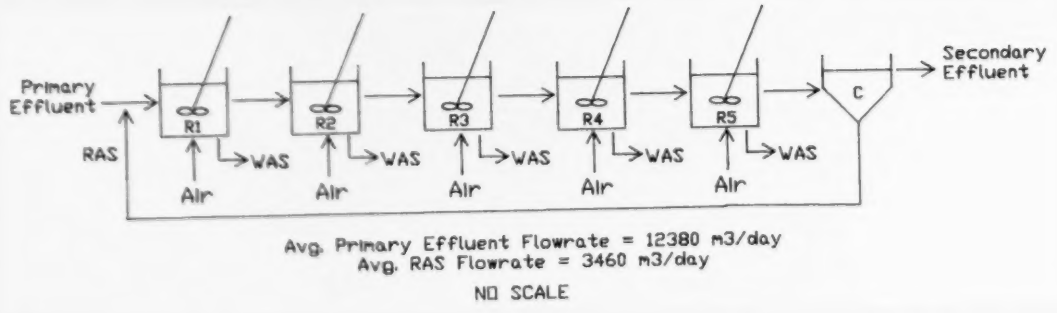

Figure 17. Jones Island Plant process flow diagram as modeled.

TABLE 19. MODEL CALIBRATION FOR THE JONES ISLAND EAST PLANT WITH FEED FRACTIONS BASED ON COD FRACTIONS

SRT = 2.8 days Temperature = 20.8 C		Measured Values	SIMULATED VALUES Default	 Default w/ Temp. adj.	 Calibrated
OUR (mg O2/L/day):	Reactor 2	1146	916.1	1032.2	1061.2
	Reactor 4	712	843.3	839.6	717.1
DO (mg-O2/L)	Reactor 2	1.6	0.2	0.3	0.0
	Reactor 4	4.9	0.9	1.9	3.3
Effluent Ammonia (mg-N/L)		5	23.3	22.7	21.1
Effluent NO3/NO2 (mg-N/L)		1	0	0.1	0.5
MLVSS (mg COD/L):	Reactor 2	2470.8	2423.4	2364.5	2584.7
MLVSS (mg COD/L):	Reactor 4	2669.6	2484.6	2410.3	2627.3
K(L)a (1/day):	Reactor 1	61	61	61	61
	Reactor 2	105	105	105	105
	Reactor 3	98.5	98.5	98.5	98.5
	Reactor 4	105	105	105	105
	Reactor 5	105	105	105	105

VALUES OF MODEL PARAMETERS FOR SIMULATIONS					
H	μ max	1/d	4.00	4.06	4.06
E	Ks COD	mg COD/L	10.0	10.2	10.2
T	Ks O2	mg O2/L	0.55		0.10
E	yield	mg cells/mg COD	0.67		
R	b decay	1/d	0.620	0.630	0.350
O	anoxic growth factor		0.8		
T	Ks NO3	mg-N/L	0.2		
R	hydrolysis rate	1/d	2.20	2.235	2.235
O	hydrol satur ratio	g COD/g COD	0.150	0.148	0.148
P	anoxic hydrol factor		0.4		
H	ammonification	L/mg COD/d	0.160	0.163	0.163
I	frac. part. prod.	mg COD/mg COD	0.080		
C	N in biomass	mg-N/mg COD	0.086		
	N in part. prod.	mg-N/mg COD	0.060		
	O2 saturation conc	mg O2/L	9.0	10.15	10.15
	μ max	1/d	0.650	0.660	0.62
A	Ks NH-N	mg-N/L	1.000		
U	Ks O2	mg O2/L	0.550		0.10
T	yield	mg cells/mg COD	0.240		
O	b decay	1/d	0.120		

simulation is realistically based on the measured K_La and DO driving force). The measured and simulated OUR profile for reactors two and four also agree reasonably well. However, for reactor two, the measured values tend to be higher and show more variation than the simulated values. This is probably a result of oxygen limitation carried over from reactor one. The measured and simulated OUR values for reactor one do not show good agreement, and this is an artifact caused by the effect of oxygen limitation on the BOD bottle uptake measurement.

South Shore Plant

Table 21 shows the process summary for the Milwaukee South Shore Plant.

TABLE 20. MODEL CALIBRATION FOR THE JONES ISLAND EAST PLANT WITH FEED FRACTIONS BASED ON BATCH REACTOR DATA

SRT = 2.8 days Temperature = 20.8 C		Measured Values	SIMULATED VALUES Default	 Default w/ Temp. adj.	 Calibrated
OUR (mg O2/L/day):	Reactor 2	1146	904.4	1015.3	1056.9
	Reactor 4	712	723.6	778.2	709.7
DO (mg-O2/L)	Reactor 2	1.6	0.3	0.4	0.1
	Reactor 4	4.9	2	2.6	3.1
Effluent Ammonia (mg-N/L)		5	19.2	16.2	12.5
Effluent NO3/NO2 (mg-n/L)		1	0.1	0.7	4.7
MLVSS (mg COD/L):	Reactor 2	2470.8	2685.9	2598.4	2887
MLVSS (mg COD/L):	Reactor 4	2669.6	2633.9	2534.5	2820.9
K(L)a (1/day):	Reactor 1	61	61	61	61
	Reactor 2	105	105	105	105
	Reactor 3	98.5	98.5	98.5	98.5
	Reactor 4	105	105	105	105
	Reactor 5	105	105	105	105

VALUES OF MODEL PARAMETERS USED FOR SIMULATIONS

H	μ max	1/d	4.00	4.06	4.06
E	Ks COD	mg COD/L	10.0	10.2	10.2
T	Ks O2	mg O2/L	0.55		0.10
E	yield	mg cells/mg COD	0.67		
R	b decay	1/d	0.620	0.630	0.230
O	anoxic growth factor		0.8		
T	Ks NO3	mg-N/L	0.2		
R	hydrolysis rate	1/d	2.20	2.235	2.235
O	hydrol satur ratio	g COD/g COD	0.150	0.148	0.148
P	anoxic hydrol factor		0.4		
H	ammonification	L/mg COD/d	0.160	0.163	0.163
I	frac. part. prod.	mg COD/mg COD	0.080		
C	N in biomass	mg-N/mg COD	0.086		
	N in part. prod.	mg-N/mg COD	0.060		
	O2 saturation conc	mg O2/L	9.0	10.15	10.15
	μ max	1/d	0.650	0.660	0.59
A	Ks NH-N	mg-N/L	1.000		
U	Ks O2	mg O2/L	0.550		0.10
T	yield	mg cells/mg COD	0.240		
O	b decay	1/d	0.120		

Figure 19 shows the configuration of the basin studied and Figure 20 shows the process flow diagram. Notable features of this study were its limited duration (11 hours) and difference in DO conditions between the basin studied (Basin 17) and the other parallel basins feeding the same group of clarifiers. These limitations make these data more difficult to interpret.

The model calibration is shown in Table 22. It was not possible to calibrate the model to reasonably simulate both the measured OUR and measured MLVSS concentrations. Simulations based on the default parameters gave OUR values much lower than the measured values, and MLVSS values somewhat lower than the measured values. Improved agreement between the simulated and measured OUR values was obtained by decreasing the heterotrophic yield and

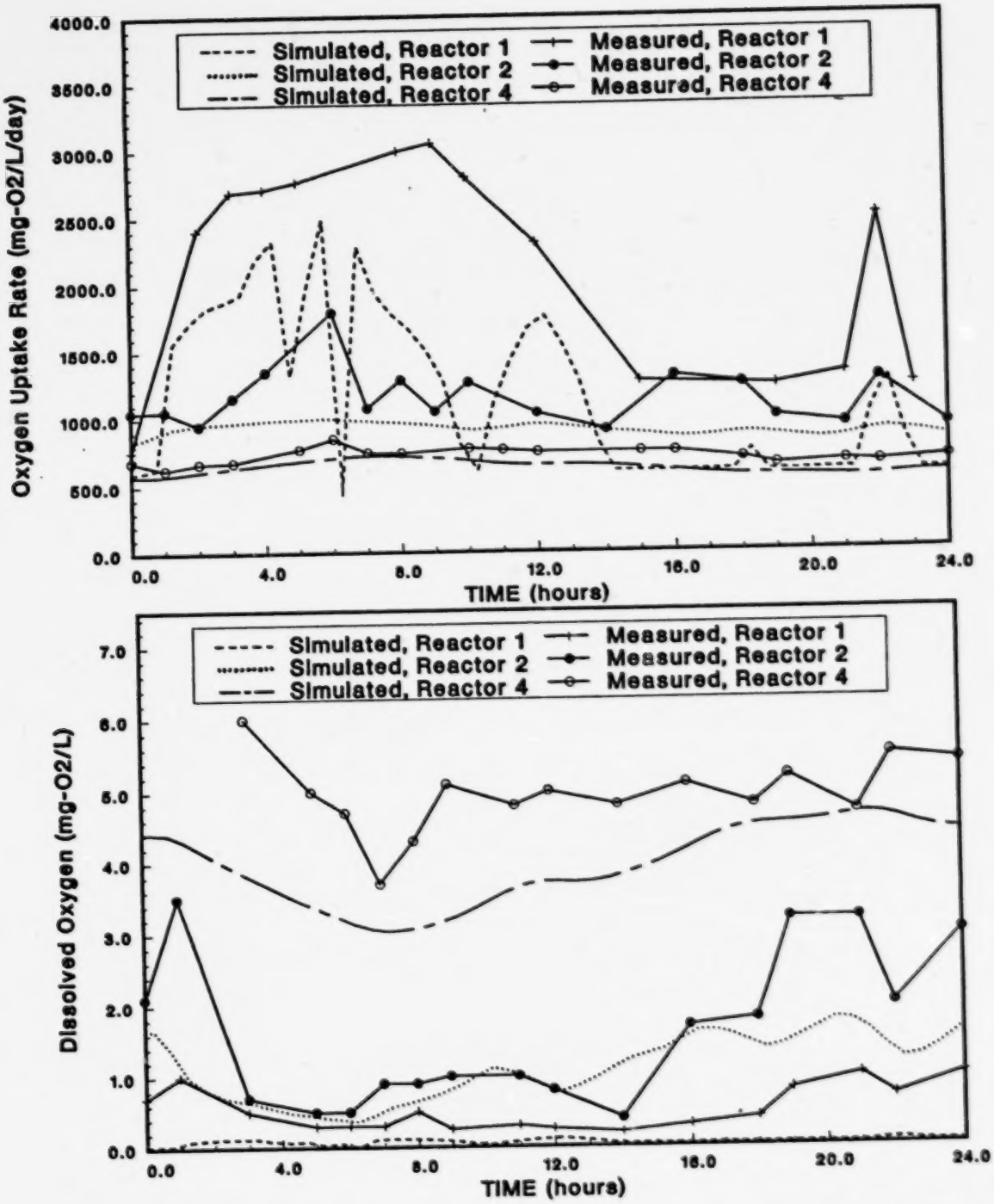

Figure 18. Jones Island Plant dynamic simulation: comparison of measured and simulated OUR and DO values.

Table 21. SOUTH SHORE PLANT PROCESS SUMMARY

Average Daily Flow: 371,000 m^3/d (98 MGD)
Average Raw Influent BOD_5: 162 mg/L
Major Industrial Contributors: glue processors, food processors, and machine industries
Fraction of Flow: 0.06
Fraction of BOD_5: 0.18
Primary Treatment: sedimentation
Nominal SRT at Time of Study: 4.3 days
Recycle Ratio at Time of Study: 0.19
Fraction of Flow Treated by Aeration Basin Studied: 0.05

Process Configuration: This plant consists of 28 single-pass basins, each 60 m. (196 ft.) long and 9.1 m. (30 ft.) wide. Mixed liquor from a group of basins is combined before clarification. There were 20 basins in operation during the study and many of these became anoxic during the night. The aeration rate and DO concentrations of Basin 17 were considerably higher than those of the other basins, and, because of this, it was selected for study midway during the study period. Thus, only 11 hours of data were taken.

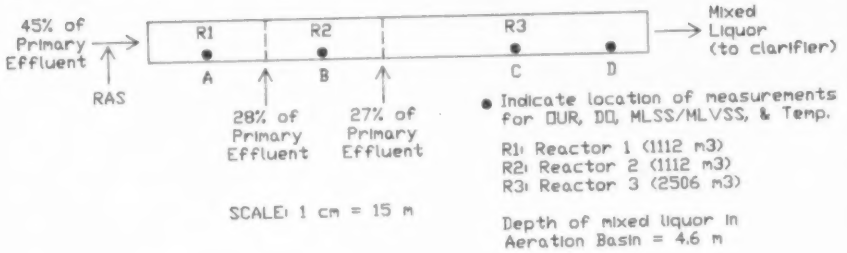

Figure 19. South Shore Plant plan sketch of basin 17.

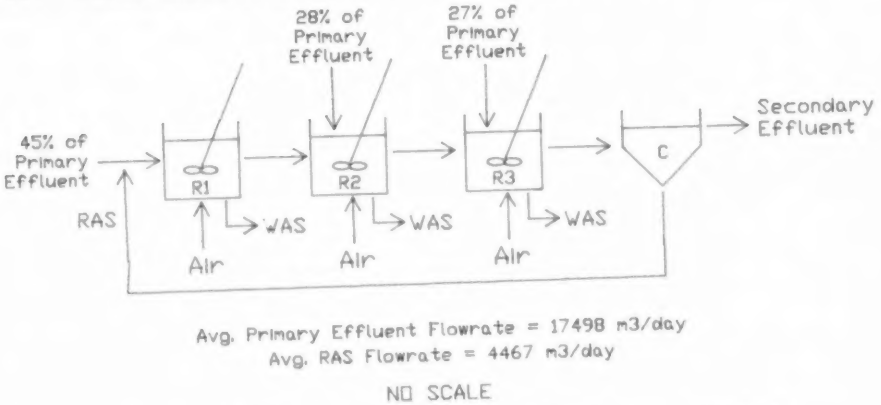

Figure 20. South Shore Plant process flow diagram as modeled.

TABLE 22. MODEL CALIBRATION FOR THE SOUTH SHORE PLANT, BASIN 17

SRT = 4.3 days Temperature = 18.5 C		Measured Values	SIMULATED VALUES Default	 Default w/ Temp. adj.	 Calibrated
OUR (mg O2/L/day):	Reactor 1	993	753.9	781.8	990.8
	Reactor 2	818	524.6	544.7	715.8
	Reactor 3	644	361.7	364.1	471.4
DO (mg-O2/L):	Reactor 1	1.1	2.5	3.8	2.1
	Reactor 2	2.1	2.5	3.7	1.8
	Reactor 3	3.5	3.6	5.2	3.6
Effluent Ammonia (mg-N/L)		3.6	1.6	1.5	2.7
Effluent NO3/NO2 (mg-N/L)		5.8	8.3	9.6	12.3
MLVSS (mg COD/L):	Reactor 1	1852	1597.9	1607.6	952.7
	Reactor 3	1505	909.9	915.6	582.1
Kla (1/day):	Reactor 1	121.0	121.0	116.8	116.8
	Reactor 2	82.0	82.0	79.1	79.1
	Reactor 3	70.0	70.0	67.6	67.6

VALUES OF MODEL PARAMETERS USED FOR SIMULATIONS

H	μ max	1/d	4.00	3.88	3.88
E	Ks COD	mg COD/L	10.0	9.7	9.7
T	Ks O2	mg O2/L	0.55		0.10
E	yield	mg cells/mg	0.67		0.46
R	b decay	1/d	0.620	0.602	0.602
O	anoxic growth factor		0.8		
T	Ks NO3	mg-N/L	0.2		
R	hydrolysis rate	1/d	2.20	2.136	2.136
O	hydrol satur ratio	g COD/g COD	0.150	0.146	0.146
P	anoxic hydrol factor		0.4		
H	ammonification	L/mg COD/d	0.160	0.155	0.155
I	frac. part. prod.	mg COD/mg CO	0.080		
C	N in biomass	mg-N/mg COD	0.086		
	N in part. prod.	mg-N/mg COD	0.060		
	O2 saturation conc	mg O2/L	9.00	10.84	10.84
	μ max	1/d	0.650	0.631	0.500
A	Ks NH-N	mg-N/L	1.000		
U	Ks O2	mg O2/L	0.550		0.100
T	yield	mg cells/mg	0.240		
O	b decay	1/d	0.120		

decay coefficients, but this further accentuated the difference between the measured and simulated MLVSS concentrations.

The South Shore Plant uses a relatively complex process control strategy in which an on-line settleometer is used to maintain the MLVSS concentration at a set value by controlling the return sludge flow rate (Grinker, 1986). Over the short period of study, this control system may have caused a change in MLVSS inventory in the clarifiers. This might explain the lack of agreement between the simulated and measured MLVSS concentrations in aeration basin 17.

TOTAL PROCESS AVERAGE OXYGEN UTILIZATION RATES

An estimate of the total daily process average oxygen utilization rate can be made by multiplying the estimated OUR values by the corresponding reactor volumes. In the absence of direct measurements of OUR values for all tank regions, the best estimates of OUR for the various reactor regions are given by the SSSP simulation based on the calibrated parameters. These estimates are consistent with the measured OUR, DO, off-gas and solids data used to calibrate the model. An example calculation for the portion of the Madison plant studied is shown in Table 23.

TABLE 23. CALCULATION OF TOTAL PROCESS AVERAGE OXYGEN UTILIZATION RATE FOR MADISON PLANT BASED ON CALIBRATED IAWPRC MODEL

Reactor	Volume m^3	OUR $g/m^3/day$	Utilization Rate kg/day
1	1,733	468.7	812.3
2	4,008	541.1	2,168.7
3	3,827	404.8	1,549.2
4	1,914	341.3	653.2
		Total	5,183.4 kg/day

Values of the best estimates of the total process average oxygen utilization rates calculated in this manner for all six plants are shown in column 2 of Table 24. These quantities apply to the portion of the plant included in the 24 hour studies. Consequently, the magnitudes of the utilization rates reflect both the size of the plant and the fraction of flow treated by the portion of the plant included in the study. Thus, the utilization rates reported for the Green Bay plant where the portion under study treated 33% of the plant flow are much greater than the rates reported for the larger Jones Island plant where the portion under study treated only 4% of the plant flow.

TABLE 24. COMPARISON OF TOTAL PROCESS AVERAGE OXYGEN UTILIZATION RATES ESTIMATED BY THE IAWPRC AND CONVENTIONAL MODELS

Plant	Oxygen Utilization Rate, kg/day				Percent Difference
	IAWPRC Model Total	Conventional Model			
		Total	R_c	R_n	
Madison	5,183	5,001	3,797	1,204	+3.5
Monroe	822	845	581	264	-2.9
Green Bay	29,350	30,600	30,448	152	-4.3
Portage Lake	593	633	592	41	-6.8
Jones Island	2,544	2,925	2,803	122	-15.
South Shore	3,079	3,532	2,352	1,180	-15.

A similar more conventional estimate may be made by viewing the process aeration tank as a single completely mixed reactor and applying the conventional model given by Equation 12 to calculate the carbonaceous oxygen utilization rate, R_C. The values of the heterotrophic yield coefficient, Y_H, used in the IAWPRC (SSSP) and conventional models are identical. However, the appropriate value of the oxidative decay coefficient, b_C, for use in the conventional model must be calculated based on the IAWPRC hydrolytic decay coefficient, b, using Equation 15. The total oxygen utilization rate is then calculated according to Equation 14 after determining the nitrogenous oxygen demand from Equation 13. An example calculation for the Madison process is shown in Table 25.

TABLE 25. CALCULATION OF TOTAL PROCESS AVERAGE OXYGEN UTILIZATION RATE FOR THE MADISON PLANT BASED ON CONVENTIONAL MODEL

From Equation 15 $b_C = b\,(1 - Y_H(1 - f_R)) = 0.626(1 - 0.600(1 - 0.2)) = 0.28\ d^{-1}$

From Equation 12

$$R_C = Q(S_{C1} - S_{C2})\left(\frac{1 + b_C\theta_C - Y_H - f_O Y_H b_C \theta_C}{1 + b_C\theta_C}\right)$$

$$R_C = 20{,}346\,(116.6 + 119.8 - 1.4)\left(\frac{1 + 0.28(16.4) - 0.60 - 0.2(0.6)0.28(16.4)}{1 + 0.28(16.4)}\right)$$

R_C = 3797 kg oxygen/day

From the definition of available nitrogen, N_{TA},
N_{TA} = Influent Nitrogen - Nitrogen in Waste Sludge
N_{TA} = 417.1 - 91.0 = 326.1 kg/day
(quantities are calculated based on measured influent data, SRT and IAWPRC simulations calibrated to measured effluent and process data)

From the definition of nitrogen nitrified to nitrate, N_{TN},
N_{TN} = Available Nitrogen - TKN in Effluent
N_{TN} = 326.1 - 20.4 = 305.7 kg/day

From the definition of nitrogen denitrified from nitrate to nitrogen gas, N_{DN},
N_{DN} = N_{TN} - Nitrate Remaining in Effluent
N_{DN} = 305.7 - 238.0 = 67.7 kg/day

From Equation 13
$R_n = 4.57\,N_{TN} - 2.86\,N_{DN}$
$R_n = 4.57\,(305.7) - 2.86\,(67.7) = 1{,}203.6$ kg/day

From Equation 14
$R_t = R_C + R_n$
$R_t = 3{,}797.6 + 1{,}203.6 = 5{,}001.2$ kg/day

Column 3 of Table 24 summarizes the total process average oxygen utilization rates calculated for all six plants using the conventional model. For most of the plants, the agreement between the IAWPRC and conventional models was excellent as the estimates generally agreed within 10%. The exceptions to this were the two Milwaukee plants, Jones Island and South Shore where the conventional approach estimated an oxygen utilization rate 15% greater than the IAWPRC approach.

It can be concluded that, for given values of the yield and decay coefficients, the two approaches produce estimates of the average process oxygen utilization rate which agree reasonably well. In this study, the IAWPRC model was calibrated to plant data and, therefore, was judged to give the better estimates of the process oxygen utilization rates. There is some indication, based on the Milwaukee plants, that the conventional approach may slightly overestimate the oxygen utilization rate.

These results reinforce the observations made earlier in Section 4, namely that the conventional approach is useful but has several disadvantages when compared with the IAWPRC model and SSSP software package. Both approaches need applicable values of the heterotrophic yield and decay coefficients. However, the conventional model does not distinguish between slowly and readily degradable substrates and cannot easily be adapted to simulate spatial and temporal variations of OUR. Moreover, the conventional approach can neither simulate nor predict nitrogenous oxygen demand. It can only calculate nitrogenous oxygen demand based on observed influent and effluent nitrogen concentrations. On the other hand, it was observed in this study that the IAWPRC model needs only one additional calibrated parameter, the maximal autotrophic specific growth rate, to simulate the entire process. In almost all cases, default values of all but two or three parameters (hetertrophic yield, hetertrophic decay, and autotrophic maximal specific growth rate) were adequate in model calibration.

SPATIAL AND TEMPORAL VARIATION IN OXYGEN UPTAKE RATES

Figure 12 shows that the IAWPRC model was able to correctly simulate the spatial variation in the temporal average DO and OUR for the Madison plant. Similar results were obtained for other plants in which several reactors were used in the simulation model. Notice that the measured OUR values in Figure 12 refer to the in-vitro values measured by the BOD bottle uptake method. These represent potential maximal values which could occur at high DO concentrations. In regions of low DO concentrations, the actual in-situ OUR values which are simulated by the model based on the off-gas K_La values are considerably lower than the potential maximal values measured by the bottle uptake method. This is because the bottle uptake method does not reflect oxygen limitations which begin to occur at DO values below approximately 1.5 mg/l.

In quantitating the spatial variation of OUR, it is useful to understand the relationship between OUR and DO and how the phenomenon of oxygen limitation smoothes the variations in OUR and distributes the oxygen utilization more uniformly over the tank volume. It was observed in studying the long, narrow aeration tanks at Madison and Jones Island that a zone of low DO and high potential maximal OUR was present in the mixed liquor distribution channel and aeration basin entrance zone. During times of high load, this zone moved into the aeration basin and progressively occupied more aeration volume until it receded under conditions of lower load. Figure 12 represents a situation in which the zone of high potential maximal OUR occupies the first third of the tank. If additional aeration capacity were provided in this

region, the DO would increase and the OUR could approach 1,200 mg/l/day, the potential maximal value, at the tank entrance. However, the faster rate of readily biodegradable COD removal would cause the OUR to decrease more rapidly with tank distance and to fall below the current, correctly simulated values. This is because the total amount of oxygen utilized would remain approximately constant. In this case, oxygen limitation smoothes the spatial variation in OUR.

From the viewpoint of aeration energy economy, it is beneficial to operate under conditions of low DO. However, because the DO concentration can influence many facets of process performance, considerations other than energy economy may also be important. The DO concentration and presence or absence of anoxic conditions are known to influence the selection of microorganisms and this, in turn, can influence biomass settleability and process efficiency.

Figures 6, 11, 15, and 18 illustrate the temporal variation in the measured and simulated spatial average OUR and DO for several plants. The IAWPRC model was able to simulate these temporal variations reasonably well. In regions of low DO, the model again correctly simulated the actual OUR significantly lower than the value measured by the BOD bottle technique.

Table 26 shows the estimated average OUR values for the various plants along with the ratios of the estimated maximal and minimal values to the average. Columns 3 to 5 describe spatial variations of the temporal daily

TABLE 26. SPATIAL AND TEMPORAL VARIATION OF OXYGEN UPTAKE RATES

Plant	Estimated Average OUR mg/l/day	Spatial Variation Ratios to Average OUR			Temporal Variation Ratios to Average OUR	
		Potential Maximum	Actual Maximum	Minimum	Maximum	Minimum
Madison	451	2.8	1.2	0.8	1.6	0.7
Monroe	454	1.1	1.1	0.9	1.1	0.8
Green Bay*	2,200	1.0	1.0	1.0	1.3	0.7
Portage Lake*	665	1.0	1.0	1.0	1.3	0.8
Jones Island	705	3.0	1.4	0.6	1.6	0.8
South Shore	651	1.5	1.4	0.6	1.1	0.9

* contact tank only

average determined from the IAWPRC model calibrated to the measured data. Columns 6 and 7 describe temporal variation of the spatial reactor average determined directly from the measured data for reactors which were not oxygen limited. It should be emphasized that these ratios are based only on the 24 hour studies conducted on the plants and represent the maximal and minimal occurrences observed during a single 24 hour period. For a longer observation period, such as a month or a year, more variation in uptake rates would be expected to occur.

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APPENDIX A

RAW DATA FROM THE 24-HOUR PLANT STUDIES

TABLE A1. RAW DATA FOR THE PORTAGE LAKE PLANT

	TIME	(PM) 12	6-15-87 1	2	3	4	5	6	7	8	9	10	11	(AM) 12	1
INFLUENT	TOTAL COD (mg/L)	325		255		448		342		349		250		287	
	SOLUBLE COD (mg/L)	140		131		253		183		207		120		135	
	AMMONIA (mg-N/L)		7.8		5		5.6		6		6		5.2		4.9
	ORGANIC-N (mg-N/L)	7.8		10.1		5		7.7		4.8		5.6		1.1	
	ALKALINITY (mol/L)	4.1												4.1	
	FLOWRATE (M3/day)	3638	3638	4374	4115	3798	3925	3611	5069	2923	3638	3638	3638	2679	3050
EFFLUENT	TOTAL COD (mg/L)	54		48		54		64		50		44		38	
	SOLUBLE COD (mg/L)	26		26		39		43		34		34		30	
	AMMONIA (mg-N/L)		2.7		3.8		3.2		3.8		3.4		3.6		2.5
	ORGANIC-N (mg-N/L)	4		1.7		1		1.2		0.5		0.7		0.8	
	NO3/NO2 (mg-N/L)		2.3		3.3		4.2		3.3		2.1		1.8		2.9
CONTACT TANK	OUR (mg O2/L/day)	528	528	595	706	626	667	695	758	706	675	600	643	602	582
	DO (mg/L)	1.8		0.5		0.5		0.5		0.5		0.7		0.8	
	MLVSS (mg/L)		1440				1450		1430		1540		1770		1980

	TIME	(AM) 2	6-16-87 3	4	5	6	7	8	9	10	(AM) 11	Average	F W Average
INFLUENT	TOTAL COD (mg/L)	171		100		200		225		250		266.8	274.9
	SOLUBLE COD (mg/L)	70		50		93		110		120		134.3	138.9
	AMMONIA (mg-N/L)		4.7		2.8		6		7.2		7	5.7	5.8
	ORGANIC-N (mg-N/L)	2.8		3.1		7.7		7.8		5.8		5.8	6.1
	ALKALINITY (mol/L)											4.1	
	FLOWRATE (M3/day)	2426	2426	2426	2876	2426	3658	4115	5778	3638	3638	3547.5	
EFFLUENT												Average	F W Average
	TOTAL COD (mg/L)	35		32		34		32		38		43.6	44.5
	SOLUBLE COD (mg/L)	26		28		30		28		33		31.4	31.6
	AMMONIA (mg-N/L)		2.2		1		0.8		1.3		2.5	2.6	2.6
	ORGANIC-N (mg-N/L)	0.3		1.6		2.3		2.5		2.6		1.6	1.7
	NO3/NO2 (mg-N/L)		6		9.6		7.5		7		4.5	4.5	4.5
CONTACT TANK												Average	
	OUR (mg O2/L/day)	600	557	577	600	551	523	513	470	527	514	597.6	
	DO (mg/L)	0.9		1.3		2.4		3.5		4		1.5	
	MLVSS (mg/L)		1950		1660		1810		1600			1663.0	

TABLE A2. RAW DATA FOR THE GREEN BAY PLANT

	TIME	(PM) 12	6-22-87 1	2	3	4	5	6	7	8	9	10	11	(AM) 12	1
INFLUENT	TOTAL COD (mg/L)	932		905		900		1105		1269		898		900	
	SOLUBLE COD (mg/L)	626		610		700		773		854		610		620	
	AMMONIA (mg/L)		32.5		32.5		36		40.5		36	40		40	
	ORGANIC-N (mg/L)	19.1		25.8		26		27.7		24.1		19.6		22.9	
	ALKALINITY (mol/L)	5												5	
	FLOWRATE (M3/day)	69303	66919	65443	60409	65632	62679	62642	59727	61847	60598	61598		58592	55375
EFFLUENT	TOTAL COD (mg/L)	280		289		298		289		250		246		253	
	SOLUBLE COD (mg/L)	176		201		220		235		221		211		190	
	AMMONIA (mg/L)		31.5		30		32		40	29	36			30	
	ORGANIC-N (mg-N/L)	3.2		4.9		3.1		2		3.9		2.8		2.8	
	NO3/NO2 (mg-N/L)														
CONTACT	OUR (mg-02/L/day)	1510	1210	1363	1426	1474	2031	1884	2124	2280	2066	2177	2117	1829	1920
	DO (mg-02/L)	2.9		3.3		1		0.7		0.8		0.9		0.5	
	MLVSS (mg/L)	1880		1960		1760		1880		1680		1960		1860	
	RAS(M3/day)	39004	39046	39061	39035	38751	38859	38796	38857	37850	38588	38633	38679	38635	38898

	TIME	(AM) 2	6-23-87 3	4	5	6	7	8	9	10	(AM) 11	Average	F W Average
INFLUENT	TOTAL COD (mg/L)	1112		1146		1300		1000		997		1038.7	1027.6
	SOLUBLE COD (mg/L)	724		895		1106		750		529		733.1	718.3
	AMMONIA (mg/L)	43		42					29			37.2	36.8
	ORGANIC-N (mg/L)	14.6		21.8		18.5		19				21.7	22.0
	ALKALINITY (mol/L)												
	FLOWRATE (M3/day)	51098	52687	47918	47313	45042	45458	53444	65821	66729	67940	58878.9	
EFFLUENT	TOTAL COD (mg/L)	293		250		263		266		263		270.0	270.7
	SOLUBLE COD (mg/L)	211		200		221		210		206		208.5	208.1
	AMMONIA (mg/L)	30	28.4	25.5	23.9		27		32.5			30.4	30.8
	ORGANIC-N (mg-N/L)	3.1		3		3.1		2.5		3.2		3.1	3.2
	NO3/NO2 (mg-N/L)												
CONTACT	OUR (mg-02/L/day)	1846	1793	2026	1762	1877	1726	1721	2040	2326	1685	1842.2	
	DO (mg-02/L)	1		1.3		1.2		2.3		1.9		1.5	
	MLVSS (mg/L)	2000		2060		2080		1960		1820		1908.3	
	RAS(M3-day)	38898	38732	38721	38766	38872	38842	38888	38974	38959	39020	38807	

TABLE A3. RAW DATA FOR THE MADISON PLANT

	TIME	(PM) 7-13-87 6	7	8	9	10	11	(AM) 7-14-87 12	1	2	3
INFLUENT	TOTAL COD (mg/L)		162		278		271		363		265
	SOLUBLE COD (mg/L)		39		191		29		65		158
	AMMONIA (mg-N/L)			20.5		41.5		10.0		10.0	
	ORGANIC-N (mg-N/L)		10.8		11.7		10.8		10.8		9.1
	FLOWRATE (M3/day)	22448	22223	22789	21580	21104	22901	19940	20459	15432	14851
EFFLUENT	TOTAL COD (mg/L)		8		16		16		16		24
	SOLUBLE COD (mg/L)		20		7		7		7		10
	AMMONIA (mg-N/L)			0.2		0.3		0.2		0.0	
	ORGANIC-N (mg-N/L)		1.0		1.9		1.7		1.3		1.3
	NO3/NO2 (mg-N/L)							15		14.6	
REACTOR (POINT B)	OUR (mg-O2/L/day)		1183	996	1084	1123	1267		1340	1250	
	DO (mg-O2/L)		0.5		0.5	0.7	0.4		0.4	0.2	
	MLVSS (mg/L)			1670.0		1600.0		1480.0		1570.0	
	RAS (M^3/day)	13569		13524		13558		13549		13535	
REACTOR (POINT E)	OUR (mg-O2/L/day)		968	913		903	996	1006		1071	982
	DO (mg-O2/L)		1.8		0.9	0.9	0.9		0.9	1.0	
	MLVSS (mg/L)		<----------			NOT MEASURED			----------		-->
REACTOR (POINT G)	OUR (mg-O2/L/day)						428			387	374
	DO (mg-O2/L)					2.2	2.2			2.4	
	MLVSS (mg/L)		<----------			NOT MEASURED			----------		-->
REACTOR (POINT G & POINT I)	OUR (mg-O2/L/day)		325	336		351	324	387		338	335
	DO (mg-O2/L)		3.9		3.8	3.6	3.9		3.7	4.0	
	MLVSS (mg/L)		<----------			NOT MEASURED			----------		-->

TABLE A3. RAW DATA FOR THE MADISON PLANT CONTINUED

	TIME	(AM) 4	5	6	7	8	9	10	11	(PM) 12	1
INFLUENT	TOTAL COD (mg/L)		245				210		167		340
	SOLUBLE COD (mg/L)		123		172		172		97		246
	AMMONIA (mg-N/L)	14.0		11.0		16.0		12.0		6.5	
	ORGANIC-N (mg-N/L)		13.2		9.9		20.7		9.9		19.2
	FLOWRATE (M3/day)	14956	12779	14371	16654	17217	22482	23076	24936	24341	23442
EFFLUENT	TOTAL COD (mg/L)		8		16		20		24		8
	SOLUBLE COD (mg/L)		20		16		23		7		7
	AMMONIA (mg-N/L)	0.2		0.3		0.4		0.3		0.1	
	ORGANIC-N (mg-N/L)		1.3		1.3		1.5		1.2		0.7
	NO3/NO2 (mg-N/L)			12.4		13.4		11.6		12.6	
REACTOR ONE (POINT B)	OUR (mg-02/L/day)	1210	1084		1115	1130		1250		1233	1858
	DO (mg-02/L)		0.3		0.3	0.3	0.5		0.6	0.3	
	MLVSS (mg/L)	1860		1500		1480		1450		1550	
	RAS (M^3/day)	13542		13533		13592		13574		13561	
REACTOR TWO (POINT E)	OUR (mg-02/L/day)		968	871	871	893		1052		1045	1115
	DO (mg-02/L)		0.8		1.1	1.0	1.4		1.4	0.8	
	MLVSS (mg/L)		<---------- NOT MEASURED ----------								
REACTOR THREE (POINT E)	OUR (mg-02/L/day)		406	324	357		319	348		450	416
	DO (mg-02/L)		3.2		3.9	4.4	4.2		3.7	3.2	
	MLVSS (mg/L)		<---------- NOT MEASURED ----------								
REACTOR FOUR (POINT G & POINT I)	OUR (mg-02/L/day)		348	335	309		329	336		374	386
	DO (mg-02/L)				4.6	4.6	4.0		4.2	3.2	
	MLVSS (mg/L)		<---------- NOT MEASURED ----------								

TABLE A3. RAW DATA FOR THE MADISON PLANT CONTINUED

		TIME	(PM) 2	3	4	5	6	Average	Flow Weighted Average
INFLUENT		TOTAL COD (mg/L)		353		246		264	263.1
		SOLUBLE COD (mg/L)		91		194		131	130.0
		AMMONIA (mg-N/L)	6.0		7.0			14.1	14.0
		ORGANIC-N (mg-N/L)		14.1		12.4		12.7	12.9
		FLOWRATE (M3/day)	24182	22992	21623	21537		20346	
EFFLUENT		TOTAL COD (mg/L)		8		16		15	15.0
		SOLUBLE COD (mg/L)		20		20		14	13.4
		AMMONIA (mg-N/L)	0.0		0.2			0.2	0.2
		ORGANIC-N (mg-N/L)		1.4		1.2		1.3	1.3
		NO3/NO2 (mg-N/L)	9.6		8.4			12.4	12.0
REACTOR ONE		OUR (mg-O2/L/day)	1711		1583		1417	1284	
	O	DO (mg-O2/L)	0.5	0.8		0.7		0.45	
	N	MLVSS (mg/L)	1500		1340			1545	
	E	RAS (M^3/day)	13540		13548			13552.0	
		(POINT B)							
REACTOR TWO		OUR (mg-O2/L/day)		1230	1226		996	1006	
	T	DO (mg-O2/L)	1.2	1.3		1.8		1.15	
	W	MLVSS (mg/L)	<---------- NOT MEASURED ---------->						
	O								
		(POINT E)							
REACTOR THREE		OUR (mg-O2/L/day)		615		739	654	447	
	T	DO (mg-O2/L)	3.1	1.8		1.5		3.0	
	H	MLVSS (mg/L)	<---------- NOT MEASURED ---------->						
	R								
	E								
	E	(POINT G)							
REACTOR FOUR		OUR (mg-O2/L/day)		448		462	441	363	
	F	DO (mg-O2/L)	3.7	3.4		3.2		3.9	
	O	MLVSS (mg/L)	<---------- NOT MEASURED ---------->						
	U								
	R								
		(POINT G & POINT I)							

TABLE A4. RAW DATA FOR THE MONROE PLANT

	TIME	(AM) 8	8-18-87 9	10	11	(PM) 12	8-19-87 1	2	3	4	5
INFLUENT	TOTAL COD (mg/L)	249		323		290		303		337	
	SOLUBLE COD (mg/L)	88		115		81		67		67	
	AMMONIA (mg-N/L)		11.0		12.0		10.0		11.3		11.0
	ORGANIC-N (mg-N/L)	21.3		18.4		18.4		19.6		22.4	
	FLOWRATE (M3/day)		2905	3278	3105	3105	3278	3278	3105	3105	3105
EFFLUENT	TOTAL COD (mg/L)	39		32		32		26		26	
	SOLUBLE COD (mg/L)	20		25		35		35		35	
	AMMONIA (mg-N/L)		0.13		0.10		0.08				0.11
	ORGANIC-N (mg-N/L)	3.1		3.1		3.9		3.6		3.4	
	NITRATE/NITRITE (mg-N/L)		13.0		12.0		11.5		11.5		11.0
REACTOR (ON)	OUR (mg-02/L/day)		404		445	419	450	457	470	457	489
	DISSOLVED OXYGEN (mg/L)		6.5		5.6			6.0			
	* MLVSS (mg/L)		1400	1300					1400		
	RAS (M^3/day)		3860	3888	3894	3924	3897	3888	3906	3906	3888
	(POINT C)										
REACTOR (TWO)	OUR (mg-02/L/day)			371	371	395	389	402	400	411	415
	DISSOLVED OXYGEN (mg/L)		6.9		7.0			6.5			
	MLVSS (mg/L)			1200					800		
	(POINT F)										

* MLVSS data from Points A, C, and D were used for this reactor.

TABLE A4. RAW DATA FOR THE MONROE PLANT CONTINUED

		(PM) 6	7	8	9	10	11	(AM) 12	1	2	3
INFLUENT	TOTAL COD (mg/L)	343		364		364		343		263	
	SOLUBLE COD (mg/L)	23		63		54		43		43	
	AMMONIA (mg-N/L)		10.5		9.2		10.0		11.0		9.7
	ORGANIC-N (mg-N/L)	23.5		23.5		24.1		23.5		25.2	
	FLOWRATE (M^3/day)	3105	3105	3105	3105	3105	3105	3105	3105	3105	3105
EFFLUENT	TOTAL COD (mg/L)	20		39		32		39		35	
	SOLUBLE COD (mg/L)	35		35		45		25		20	
	AMMONIA (mg-N/L)		0.08		0.10		0.10		0.10		0.12
	ORGANIC-N (mg-N/L)	3.1		4.2		3.9		3.9		3.1	
	NITRATE/NITRITE (mg-N/L)		14.0		13.5		10.5		11.0		11.3
REACTOR (POINT C)	OUR (mg-O2/L/day)	536	485	509	485		452	477		489	521
	DISSOLVED OXYGEN (mg/L)	5.5		6.2			6.6			6.3	
	* MLVSS (mg/L)					900	1400				
	RAS (M^3/day)	3906	3888	3888	3888	3888	3888	3888	3888	3888	3888
REACTOR (POINT F)	OUR (mg-O2/L/day)		456	480	449	402	414	441		432	433
	DISSOLVED OXYGEN (mg/L)	6.0		6.6			6.5			6.6	
	MLVSS (mg/L)						1300				

* MLVSS data from Points A, C, and D were used for this reactor.

TABLE A4. RAW DATA FOR THE MONROE PLANT CONTINUED

		(AM) 4	5	6	7	8	Average	Flow Weighted Average
INFLUENT	TOTAL COD (mg/L)	263		188			303	319.2
	SOLUBLE COD (mg/L)	60		53			63	61.9
	AMMONIA (mg-N/L)		9.5		11.7		10.6	10.6
	ORGANIC-N (mg-N/L)	23.5		20.2			22.0	22.2
	FLOWRATE (M3/day)	3105	3105		3016		3115	
EFFLUENT	TOTAL COD (mg/L)	35		25		30	32	31.6
	SOLUBLE COD (mg/L)	15		15		20	28	30.5
	AMMONIA (mg-N/L)		0.11		0.12		0.10	
	ORGANIC-N (mg-N/L)	3.6		4.2		5.0	3.7	3.6
	NITRATE/NITRITE (mg-N/L)		10.0		8.2		11.5	11.5
REACTOR (ON)	OUR (mg-O2/L/day)	481	494	495	543	489	478	
	DISSOLVED OXYGEN (mg/L)		6.2			6.3	6.1	
	* MLVSS (mg/L)	1300					1283	
	RAS (M^3/DAY)	3888	3888	3888	3888	3888	3891	
	(POINT C)							
REACTOR (TO WO)	OUR (mg-O2/L/day)	409	419	449	419	434	418	
	DISSOLVED OXYGEN (mg/L)		6.5			6.0	6.5	
	MLVSS (mg/L)	900				900	1050	
	(POINT F)							

* MLVSS data from Points A, C, and D were used for this reactor.

TABLE A5. RAW DATA FOR JONES ISLAND EAST PLANT

			(AM)		(PM)	7-28-87						
		TIME	10	11	12	1	2	3	4	5	6	7
I												
N		TOTAL COD (mg/L)	439		530				638		530	
F		SOLUBLE COD (mg/L)	276		474				558		420	
L		AMMONIA (mg-N/L)		17		16		13.5		14		20
U		ORGANIC-N (mg-N/L)	29.1		30.8		26.5		33.6		22.4	
E		FLOWRATE (M3/day)	11130	15290	15100	14530	14720	14530	14530	13210	13970	14160
N												
T												
E												
F		TOTAL COD (mg/L)	22		41		58		53		53	
F		SOLUBLE COD (mg/L)	19		28		50		39		33	
L		AMMONIA (mg/L)		5		4.6		5.7		7.4		8
U		ORGANIC-N (mg-N/L)	3.08		3.64		3.92		5.32		4.48	
E		NO3/NO2 (mg-N/L)	1.5				1		1.3			
N												
T												
R		OUR (mg O2/L/day)	(759)		2400	2688	2711	2765			2995	3050
E		DO (mg-O2/L)	0.7	1		0.5		0.3	0.3	0.3	0.5	0.25
A	O	MLVSS (mg/L)			1540		2112			1722		
C	N	RAS (M^3/DAY)	----------ONLY AVERAGE DAILY FLOW WAS DETERMINED----------									
T	E											
O		(POINT A)										
R												
R		OUR (mg O2/L/day)	1047	1053	946	1152	1344		1792	1072	1280	1047
E		DO (mg-O2/L)	2.1	3.5		0.7		0.5	0.5	0.9	0.9	1
A	T	MLVSS (mg/L)	----------NOT MEASURED----------									
C	W											
T	O											
O		(POINT B)										
R												
R		OUR (mg O2/L/day)	683	623	665	672		768	845	741	737	
E	F	DO (mg-O2/L)				6		5	4.7	3.7	4.3	5.1
A	O	MLVSS (mg/L)				1880 (ONLY MEASUREMENT)						
C	U											
T	R											
O		(POINT C)										
R												

TABLE A5. RAW DATA FOR JONES ISLAND EAST PLANT CONTINUED

	TIME	(PM) 8	9	10	11	(AM) 12	7-29-87 1	2	3	4	5
INFLUENT	TOTAL COD (mg/L)	439		547		496		519		639	
	SOLUBLE COD (mg/L)	342		332		402		444		323	
	AMMONIA (mg-N/L)		12						14.5		14
	ORGANIC-N (mg-N/L)	24.1		39.1		52.3		53.2		39.8	
	FLOWRATE (M3/day)	13590		13970		11700	11130	9420	9240	9240	8860
EFFLUENT	TOTAL COD (mg/L)	50		61		61		50		53	
	SOLUBLE COD (mg/L)	53		47		55				53	
	AMMONIA (mg/L)		3.6			4.5			4.9		4.3
	ORGANIC-N (mg-N/L)	4.2		4.48		4.48		3.92		3.92	
	NO3/NO2 (mg-N/L)	0.8		2		0.8			0.7		0.8
REACTOR O	OUR (mg O2/L/day)	2798		2304			1267			1234	
	DO (mg-O2/L)		0.3	0.25		0.2		0.3		0.4	0.8
	MLVSS (mg/L)	1570				1728			1912		
	RAS (M^3/DAY)	----------ONLY AVERAGE DAILY FLOW WAS DETERMINED----------									
	(POINT A)										
REACTOR TWO	OUR (mg O2/L/day)	1260		1029		905		1306		1248	998
	DO (mg-O2/L)		1	0.8		0.4		1.7		1.8	3.2
	MLVSS (mg/L)	----------NOT MEASURED----------									
	(POINT B)										
REACTOR FOUR	OUR (mg O2/L/day)	768	757	741			743	743		691	640
	DO (mg-O2/L)		4.8	5		4.8		5.1		4.8	5.2
	MLVSS (mg/L)										
	(POINT C)										

TABLE A5. RAW DATA FOR JONES ISLAND EAST PLANT CONTINUED

	TIME	(AM) 6	7	8	9	10	Average	Flow Weighted Average
INFLUENT	TOTAL COD (mg/L)	512		611		505	534	532.8
	SOLUBLE COD (mg/L)	246		330		309	377	378.5
	AMMONIA (mg-N/L)		25			17	16	16.7
	ORGANIC-N (mg-N/L)	31.9		30.8			35	30.6
	FLOWRATE (M3/day)	9046	9800	10750		14530	12384	
EFFLUENT	TOTAL COD (mg/L)	65		47		71	53	52.9
	SOLUBLE COD (mg/L)	59		41		53	44	41.1
	AMMONIA (mg/L)		3.8		3.7		5	5.3
	ORGANIC-N (mg-N/L)	3.92		4.61		3.78	4.1	4.2
	NO3/NO2 (mg-N/L)		0.7		0.8		1	1.0
REACTOR (POINT A)	OUR (mg O2/L/day)		1324	2496	1234		2145	
	DO (mg-O2/L)		1	0.7		1	0.5	
	MLVSS (mg/L)					1598	1740	
	RAS (M^3/DAY)						3460	
REACTOR (POINT B)	OUR (mg O2/L/day)		942	1280		929	1146	
	DO (mg-O2/L)		3.2	2		3	1.6	
	MLVSS (mg/L)		---------NOT MEASURED--------					
REACTOR (POINT C)	OUR (mg O2/L/day)		665	653		683	712	
	DO (mg-O2/L)		4.7	5.5		5.4	4.9	
	MLVSS (mg/L)						1880	

TABLE A6. RAW DATA FOR THE SOUTH SHORE PLANT BASIN 17

		(AM)	7-30-87		(PM)					
	TIME	9	10	11	12	1	2	3	4	5
I N F L U E N T	TOTAL COD (mg/L)	201		217		223		201		214
	SOLUBLE COD (mg/L)	121		149		121		103		126
	AMMONIA (mg-N/L)		15.5		26		29		34	
	TOTAL ORG-N (mg-N/L)	6.72		7.28		9.52		8.96		8.96
	FLOWRATE (m3/day)									
	BASIN # 17 WAS NOT MONITORED UNTIL AFTER 10:00 PM									
E F F L U E N T	TOTAL COD (mg/L)	35		41		50		18		38
	SOLUBLE COD (mg/L)	20		14		32		14		20
	AMMONIA (mg-N/L)		2.8			4.5	2.7		2.2	
	TOTAL ORG-N (mg-N/L)	2.24		3.92		2.8		1.96		2.8
	NO3/NO2 (mg-N/L)		5			7	7		6.5	
R E A C T O R (POINT A)	OUR (mg O2/L/day)									
	DO (mg-O2/L)									
	MLVSS (mg/L)									
	RAS (M^3/DAY)									
R E A C T O R (POINT B)	OUR (mg O2/L/day)									
	DO (mg-O2/L)									
	MLVSS (mg/L)	---------- NOT MEASURED ----------								
R E A C T O R (POINT C AND D)	OUR (mg O2/L/day)									
	DO (mg-O2/L)									
	MLVSS (mg/L)									

TABLE A6. RAW DATA FOR THE SOUTH SHORE PLANT BASIN 17 CONTINUED

			(PM)					(AM)	7-31-87		
	TIME		6	7	8	9	10	11	12	1	2	3
INFLUENT	TOTAL COD (mg/L)			175		210		243		227		207
	SOLUBLE COD (mg/L)			69		103		132		109		143
	AMMONIA (mg-N/L)		31		36							
	TOTAL ORG-N (mg-N/L)			6.72		10.6		8.4		7.84		6.16
	FLOWRATE (m3/day)	R 1:						8274	8524	8421	8395	8346
		R 2:						5102	5212	5193	5142	5114
		R 3:						5102	5212	5193	5142	5114
		TOTAL:						18478	18948	18807	18679	18574
EFFLUENT	TOTAL COD (mg/L)			29		29		29		32		17
	SOLUBLE COD (mg/L)			14		20		26		17		17
	AMMONIA (mg-N/L)		2.1		3.4							
	TOTAL ORG-N (mg-N/L)			1.96		2.24		2.24		2.52		4.98
	NO3/NO2 (mg-N/L)		5		1.5		3		7.5		5.5	
REACTOR ONE	OUR (mg O2/L/day)						1029				960	995
	DO (mg-O2/L)							1.8		1	0.8	
	MLVSS (mg/L)											
	RAS (M^3/DAY)											
	(POINT A)											
REACTOR TWO	OUR (mg O2/L/day)											
	DO (mg-O2/L)							2.8				
	MLVSS (mg/L)		----------					NOT MEASURED				----------
	(POINT B)											
REACTOR THREE	OUR (mg O2/L/day)							662		683	620	
	DO (mg-O2/L)							4.7		3.2	3.3	
	MLVSS (mg/L)							1068				
	(POINT C AND D)											

TABLE A6. RAW DATA FOR THE SOUTH SHORE PLANT BASIN 17 CONTINUED

	TIME		(AM) 7-31-87 5	6	7	8	9	Average	Flow Weighted Average
INFLUENT	TOTAL COD (mg/L)		148		148		128	195.5	186.9
	SOLUBLE COD (mg/L)		140		122		87	117.3	123.4
	AMMONIA (mg-N/L)			11.2		21.3		25.5	
	TOTAL ORG-N (mg-N/L)		8.4		8.4		7.84	8.1	7.8
	FLOWRATE (m3/day)	R 1:	7948	7646	7233	6942	6442	7847.6	
		R 2:	4897	4694	4454	4289	3976	4825.4	
		R 3:	4897	4694	4454	4289	3976	4825.4	
		TOTAL:	17742	17034	16141	15520	14394	17498.4	
EFFLUENT	TOTAL COD (mg/L)		28.5				17	30.3	25.0
	SOLUBLE COD (mg/L)		17					19.2	19.3
	AMMONIA (mg-N/L)				5.04	5.6		3.5	
	TOTAL ORG-N (mg-N/L)		3.08		1.68			2.7	2.9
	NO3/NO2 (mg-N/L)		5	5		6		5.3	5.8
REACTOR (POINT A)	OUR (mg O2/L/day)		988		1005		981	993	
	DISSOLVED OXYGEN (mg O2/L)		1			1	1.2	1.1	
	MLVSS (mg/L)						1304	1304	
	RAS (M^3/DAY)							4444	
REACTOR (POINT B)	OUR (mg O2/L/day)							818	
	DISSOLVED OXYGEN (mg O2/L)		1.8			2	2.3	2.1	
	MLVSS (mg/L)		----------NOT MEASURED----------						
REACTOR (POINT C AND D)	OUR (mg O2/L/day)		638		648		614	643.9	
	DISSOLVED OXYGEN (mg O2/L)		3			3.1	3.9	3.5	
	MLVSS (mg/L)		1052					1060.0	

APPENDIX B

STEADY-STATE SIMULATIONS

TABLE B.1. PORTAGE LAKE PLANT SSSP STEADY STATE SIMULATION WITH CALIBRATED PARAMETERS

PROCESS CONFIGURATION AND FLOW DISTRIBUTION

OVERALL PLANT SPECIFICATIONS:

Number of Reactors (up to 9)	=	2
Solids Retention Time (days)	=	10.6
Average Flow Rate (m3/day)	=	3547

INDIVIDUAL REACTOR SPECIFICATIONS:		1	2
Reactor Volume (m3)	=	530	350
Feed Fraction (0 to 1)	=	0.00	1.00
Mass Transfer Coeff for O2 (day-1)	=	72.0	72.0
Recycle Input (m3/day)	=	3653	0
Recirculation Input (m3/day)	=	0	0
Recirculation originated from reactor	=	*	*

STEADY-STATE SOLUTION WITH FEED FRACTIONATION BASED ON TOTAL & SOLUBLE COD

CONSTITUENTS			FEED	1	2
Heterotrophic Organisms	g cod m-3	=	0.0	1466.9	766.9
Autotrophic Organisms	g cod m-3	=	0.0	26.5	13.4
Particulate Products	g cod m-3	=	0.0	1392.8	705.1
Inert Particulates	g cod m-3	=	13.6	731.1	375.2
Particulate Organics	g cod m-3	=	122.4	496.8	294.9
Soluble Organics	g cod m-3	=	107.3	3.3	9.1
Soluble Ammonia N	g n m-3	=	5.8	0.5	0.7
Soluble Nitrate/Nitrite N	g n m-3	=	0.0	2.8	1.6
Soluble Organic N	g n m-3	=	2.2	0.6	0.7
Biodegrad Part Organic N	g n m-3	=	2.5	35.0	18.8
Oxygen	g o2 m-3	=	0.0	1.2	1.5
Alkalinity	mole m-3	=	5.0	4.4	4.5
MLVSS	g cod m-3	=		4114.1	2155.5
O2 Consumed	g o2 m-3 d-1	=		695.0	662.8
Nitrate Consumed	g no3-n m-3 d-1	=		12.8	9.8

TABLE B.2. GREEN BAY PLANT SSSP STEADY-STATE SIMULATION WITH CALIBRATED PARAMETERS

PROCESS CONFIGURATION AND FLOW DISTRIBUTION

OVERALL PLANT SPECIFICATIONS:

Number of Reactors (up to 9)	= 2
Solids Retention Time (days)	= 3.1
Average Flow Rate (m3/day)	= 58879

INDIVIDUAL REACTOR SPECIFICATIONS:		1	2
Reactor Volume (m3)	=	5607	10443
Feed Fraction (0 to 1)	=	0.00	1.00
Mass Transfer Coeff for O2 (day-1)	=	118.0	311.0
Recycle Input (m3/day)	=	38840	0
Recirculation Input (m3/day)	=	0	0
Recirculation originated from reactor	=	*	*

STEADY-STATE SOLUTION WITH FEED FRACTIONATION BASED ON TOTAL & SOLUBLE COD

CONSTITUENTS			FEED	1	2
Particulate Products	g cod m-3	=	0.0	678.7	272.1
Inert Particulates	g cod m-3	=	30.9	567.3	238.0
Particulate Organics	g cod m-3	=	278.4	704.8	307.0
Soluble Organics	g cod m-3	=	510.2	2.6	7.9
Soluble Ammonia N	g n m-3	=	36.7	39.6	32.2
Soluble Nitrate/Nitrite N	g n m-3	=	0.0	0.0	0.0
Soluble Organic N	g n m-3	=	12.2	0.3	0.7
Biodegrad Part Organic N	g n m-3	=	6.7	48.2	18.1
Oxygen	g o2 m-3	=	0.0	0.1	2.5
Alkalinity	mole m-3	=	3.9	4.1	3.6
MLVSS	g cod m-3	=		5670.2	2445.8
O2 Consumed	g o2 m-3 d-1	=		1133.1	2201.5
Nitrate Consumed	g no3-n m-3 d-1	=		0.0	0.0

STEADY-STATE SOLUTION WITH FEED FRACTIONATION BASED ON BATCH REACTOR DATA

CONSTITUENTS			FEED	1	2
Particulate Products	g cod m-3	=	0.0	679.1	272.2
Inert Particulates	g cod m-3	=	30.9	567.3	238.0
Particulate Organics	g cod m-3	=	536.0	681.0	299.0
Soluble Organics	g cod m-3	=	253.0	4.9	7.9
Soluble Ammonia N	g n m-3	=	36.8	37.7	32.7
Soluble Nitrate/Nitrite N	g n m-3	=	0.0	0.0	0.0
Soluble Organic N	g n m-3	=	6.1	0.2	0.7
Biodegrad Part Organic N	g n m-3	=	12.8	41.8	14.4
Oxygen	g o2 m-3	=	0.0	0.0	2.5
Alkalinity	mole m-3	=	3.9	3.9	3.6
MLVSS	g cod m-3	=		5648.9	2438.7
O2 Consumed	g o2 m-3 d-1	=		1139.5	2201.2
Nitrate Consumed	g no3-n m-3 d-1	=		0.0	0.0

TABLE B.3. MADISON PLANT SSSP STEADY-STATE SIMULATION WITH CALIBRATED PARAMETERS

PROCESS CONFIGURATION AND FLOW DISTRIBUTION

OVERALL PLANT SPECIFICATIONS:

Number of Reactors (up to 9) = 4
Solids Retention Time (days) = 16.4
Average Flow Rate (m3/day) = 20346

INDIVIDUAL REACTOR SPECIFICATIONS:		1	2	3	4
Reactor Volume (m3)	=	1733	4008	3827	1914
Feed Fraction (0 to 1)	=	1.00	0.00	0.00	0.00
Mass Transfer Coeff for O2 (day-1)	=	45.0	68.0	50.0	51.0
Recycle Input (m3/day)	=	13552	0	0	0
Recirculation Input (m3/day)	=	0	0	0	0
Recirculation originated from reactor	=	*	*	*	*

STEADY-STATE SOLUTION WITH FEED FRACTIONATION BASED ON TOTAL & SOLUBLE COD

CONSTITUENTS			FEED	1	2	3	4
Heterotrophic Organisms	g cod m-3	=	0.0	668.0	671.3	663.1	657.5
Autotrophic Organisms	g cod m-3	=	0.0	46.5	47.2	47.5	47.6
Particulate Products	g cod m-3	=	0.0	544.3	548.4	552.2	554.1
Inert Particulates	g cod m-3	=	13.3	386.0	386.0	386.0	386.0
Particulate Organics	g cod m-3	=	119.8	130.0	93.3	69.6	60.2
Soluble Organics	g cod m-3	=	116.6	11.7	2.0	1.5	1.4
Soluble Ammonia N	g n m-3	=	14.0	7.7	3.4	1.2	0.6
Soluble Nitrate/Nitrite N	g n m-3	=	0.0	0.2	5.6	9.3	10.7
Soluble Organic N	g n m-3	=	5.5	0.7	0.5	0.4	0.4
Biodegrad Part Organic N	g n m-3	=	5.6	8.4	6.6	5.3	4.8
Oxygen	g o2 m-3	=	0.0	0.1	2.3	2.5	3.5
Alkalinity	mole m-3	=	3.9	3.4	2.7	2.3	2.2
MLVSS	g cod m-3	=		1774.8	1746.2	1718.4	1705.4
O2 Consumed	g o2 m-3 d-1	=		469.2	537.9	400.0	340.4
Nitrate Consumed	g no3-n m-3 d-1	=		100.0	3.8	2.7	1.8

TABLE B.4. MONROE PLANT SSSP STEADY-STATE SIMULATION WITH CALIBRATED PARAMETERS

PROCESS CONFIGURATION AND FLOW DISTRIBUTION

OVERALL PLANT SPECIFICATIONS:

Number of Reactors (up to 9)	=	2
Solids Retention Time (days)	=	8.4
Average Flow Rate (m3/day)	=	3115

INDIVIDUAL REACTOR SPECIFICATIONS:		1	2
Reactor Volume (m3)	=	852	996
Feed Fraction (0 to 1)	=	0.90	0.10
Mass Transfer Coeff for O2 (day-1)	=	155.0	115.0
Recycle Input (m3/day)	=	3891	0
Recirculation Input (m3/day)	=	0	0
Recirculation originated from reactor	=	*	*

STEADY-STATE SOLUTION WITH FEED FRACTIONATION BASED ON TOTAL & SOLUBLE COD

CONSTITUENTS			FEED	1	2
Heterotrophic Organisms	g cod m-3	=	0.0	686.3	658.4
Autotrophic Organisms	g cod m-3	=	0.0	31.7	30.7
Particulate Products	g cod m-3	=	0.0	372.2	361.5
Inert Particulates	g cod m-3	=	25.7	375.4	359.7
Particulate Organics	g cod m-3	=	231.6	145.2	103.4
Soluble Organics	g cod m-3	=	31.4	3.4	2.6
Soluble Ammonia N	g n m-3	=	10.6	3.3	2.0
Soluble Nitrate/Nitrite N	g n m-3	=	0.0	14.4	17.7
Soluble Organic N	g n m-3	=	2.0	0.7	0.6
Biodegrad Part Organic N	g n m-3	=	14.9	11.0	8.2
Oxygen	g o2 m-3	=	0.0	6.4	6.4
Alkalinity	mole m-3	=	5.0	3.4	3.1
MLVSS	g cod m-3	=		1610.8	1513.8
O2 Consumed	g o2 m-3 d-1	=		486.7	389.9
Nitrate Consumed	g no3-n m-3 d-1	=		1.5	1.2

TABLE B.5 JONES ISLAND EAST PLANT SSSP STEADY-STATE SIMULATION WITH CALIBRATED PARAMETERS

PROCESS CONFIGURATION AND FLOW DISTRIBUTION

OVERALL PLANT SPECIFICATIONS:

Number of Reactors (up to 9) =	5
Solids Retention Time (days) =	2.8
Average Flow Rate (m3/day) =	12380

INDIVIDUAL REACTOR SPECIFICATIONS:		1	2	3	4	5
Reactor Volume (m3)	=	240	601	601	1203	962
Feed Fraction (0 to 1)	=	1.00	0.00	0.00	0.00	0.00
Mass Transfer Coeff for O2 (day-1)	=	61.0	105.0	98.5	105.0	103.0
Recycle Input (m3/day)	=	3460	0	0	0	0
Recirculation Input (m3/day)	=	0	0	0	0	0
Recirculation originated from reactor	=	*	*	*	*	*

STEADY-STATE SOLUTION WITH FEED FRACTIONATION BASED ON TOTAL & SOLUBLE COD

CONSTITUENTS		FEED	1	2	3	4	5
Particulate Products	g cod m-3 =	0.0	158.1	160.3	162.6	167.3	171.2
Inert Particulates	g cod m-3 =	15.4	148.0	148.0	148.0	148.0	148.0
Particulate Organics	g cod m-3 =	138.9	226.5	230.7	234.1	163.7	123.1
Soluble Organics	g cod m-3 =	337.4	239.7	138.2	44.4	2.0	1.2
Soluble Ammonia N	g n m-3 =	16.7	28.3	24.9	20.0	19.0	20.4
Soluble Nitrate/Nitrite N	g n m-3 =	0.0	0.0	0.0	0.0	0.2	0.4
Soluble Organic N	g n m-3 =	18.8	2.6	0.3	0.1	0.3	0.3
Biodegrad Part Organic N	g n m-3 =	7.7	15.3	16.1	16.8	12.2	9.6
Oxygen	g o2 m-3 =	0.0	0.0	0.0	0.0	3.0	5.3
Alkalinity	mole m-3 =	5.0	5.8	5.6	5.2	5.2	5.2
MLVSS	g cod m-3 =		2523.3	2584.4	2639.7	2627.0	2599.5
O2 Consumed	g o2 m-3 d-1 =		616.7	1061.2	995.7	715.3	459.6
Nitrate Consumed	g no3-n m-3 d-1 =		6.9	1.5	1.5	3.0	1.6

STEADY-STATE SOLUTION WITH FEED FRACTIONATION BASED ON BATCH REACTOR DATA

		FEED	1	2	3	4	5
Particulate Products	g cod m-3 =	0.0	114.5	116.1	117.8	121.2	124.0
Inert Particulates	g cod m-3 =	15.4	148.0	148.0	148.0	148.0	148.0
Particulate Organics	g cod m-3 =	325.0	368.5	344.5	279.3	180.3	124.5
Soluble Organics	g cod m-3 =	151.0	88.9	11.1	2.2	1.5	1.1
Soluble Ammonia N	g n m-3 =	16.7	19.5	15.5	14.3	13.0	12.1
Soluble Nitrate/Nitrite N	g n m-3 =	0.0	0.0	0.0	0.5	2.5	4.4
Soluble Organic N	g n m-3 =	8.4	1.1	0.2	0.4	0.3	0.2
Biodegrad Part Organic N	g n m-3 =	18.1	22.3	21.3	17.7	12.0	8.7
Oxygen	g o2 m-3 =	0.0	0.0	0.1	1.0	3.2	4.3
Alkalinity	mole m-3 =	5.0	5.2	4.9	4.8	4.6	4.4
MLVSS	g cod m-3 =		2847.9	2886.1	2864.5	2819.9	2791.5
O2 Consumed	g o2 m-3 d-1 =		616.8	1056.9	871.6	702.1	581.1
Nitrate Consumed	g no3-n m-3 d-1 =		67.4	13.1	14.6	4.6	2.8

TABLE B.6. SOUTH SHORE PLANT SSSP STEADY-STATE SIMULATION WITH CALIBRATED PARAMETERS

PROCESS CONFIGURATION AND FLOW DISTRIBUTION

OVERALL PLANT SPECIFICATIONS:

Number of Reactors (up to 9)	= 3
Solids Retention Time (days)	= 4.3
Average Flow Rate (m3/day)	= 17498

INDIVIDUAL REACTOR SPECIFICATIONS:		1	2	3
Reactor Volume (m3)	=	1112	1112	2506
Feed Fraction (0 to 1)	=	0.45	0.28	0.28
Mass Transfer Coeff for O2 (day-1)	=	121.0	82.0	70.0
Recycle Input (m3/day)	=	4467	0	0
Recirculation Input (m3/day)	=	0	0	0
Recirculation originated from reactor	=	*	*	*

STEADY-STATE SOLUTION WITH FEED FRACTIONATION BASED ON TOTAL & SOLUBLE COD

CONSTITUENTS			FEED	1	2	3
Heterotrophic Organisms	g cod m-3	=	0.0	620.0	450.8	353.6
Autotrophic Organisms	g cod m-3	=	0.0	46.6	33.7	26.8
Particulate Products	g cod m-3	=	0.0	117.2	84.9	67.6
Inert Particulates	g cod m-3	=	6.3	128.7	93.4	73.7
Particulate Organics	g cod m-3	=	57.1	93.0	68.8	48.7
Soluble Organics	g cod m-3	=	104.1	3.4	3.1	2.2
Soluble Ammonia N	g n m-3	=	15.0	4.2	4.0	2.6
Soluble Nitrate/Nitrite N	g n m-3	=	0.0	10.5	10.6	12.5
Soluble Organic N	g n m-3	=	3.5	0.6	0.6	0.5
Biodegrad Part Organic N	g n m-3	=	1.9	5.7	4.2	3.1
Oxygen	g o2 m-3	=	0.0	2.2	2.1	3.9
Alkalinity	mole m-3	=	5.0	3.5	3.5	3.2
MLVSS	g cod m-3	=		1005.5	731.6	570.4
O2 Consumed	g o2 m-3 d-1	=		1023.6	707.9	464.7
Nitrate Consumed	g no3-n m-3 d-1	=		8.8	6.2	2.1